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(NASA-CR-161655) COAL GASIFICATION SYSTEMS
ENGINEERING AND ANALYSIS. APPENDIX A: COAL
GASIFICATION CATALOG Final Report (BDM
Corp., Huntsville, Ala.) 221 p
HC A10/MF A01

N81-18214

Unclas

CSCL 21D G3/28 41562





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COAL GASIFICATION SYSTEMS
ENGINEERING AND ANALYSIS
FINAL REPORT
APPENDIX A - COAL GASIFICATION CATALOG
December 31, 1980

BDM/H-80-800-TR

This Technical Report is submitted to George C. Marshall Space Flight Center under Contract Number NAS8-33824.

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1.0 INTRODUCTION

COAL GASIFICATION DATA CATALOG

1.0 INTRODUCTION

1.1 Task 5.1

This task develops the basis for further study and ultimately, the A/E evaluation and recommended facility designs of Task 5.5. Definition of candidate systems/subsystems includes the identity of and alternates for each process unit, raw material requirements, and the cost and design drivers for each process design. Alternate products studies and a marketing analysis of by-products have been performed.

1.2 Output

The output of Task 5.1 is the Coal Gasification Data Catalog contained herein. Accompanying this document in separate covers are two facility reference documents (preliminary) based on Texaco and Koppers-Totzek gasifier technologies. The preliminary reference facility documents will be revised in Task 5.3 to accommodate the results of Tasks 5.2.

1.3 Scope of Work

The scope of work in preparing the Coal Gasification Data Catalog included the following subtasks:

- Candidate System Subsystem Definition.
- Raw Materials Analysis.
- Market Analysis for By-Products.
- Alternate Products Analysis.
- Preliminary Integrated Facility Requirements.

1.3.1 Candidate System/Subsystem Definition - Task 5.1.1

Definition of candidate systems/subsystems involves the characterization of each facility subsystem constituting the integrated facility and each coal gasification module in terms of parameters required to support conceptual design costing, cost and performance sensitivity analysis, raw material and by-product analysis, coal variation and utility trades, and supporting technology requirements.

The characterization of each system/subsystem element includes the identification of major components and the design criteria which have major impact on design or cost of the components.

These data have been developed from the subsystem designs and trade-off studies previously completed in other studies by the BDM-Mittelhauser Team as well as from other published documents and proprietary studies presently in the Teams' libraries.

The subsystem characterization data have been summarized from appropriate flowsheets. Data presented in the form of detailed matrices include major components, long lead time components, inputs, outputs, by-products, raw materials, and typical operating conditions. In addition, critical technology issues are identified.

Major cost elements of the system/subsystem have been defined and include such items as:

- Capital cost.
- System capacity.
- Operating costs (manpower and chemicals and catalyst).
- Maintenance costs.

1.3.2 Raw Material Analysis - Task 5.1.2

An analysis of raw material requirements has been performed for each system that has been characterized to identify the type, quantity, quality, etc., of raw material (other than coal) required to support the TVA Coal Gasification Facility.

These raw material requirements have been described as to:

- Identity of raw material.
- Quantity or consumption of each raw material.
- Source.
- Costs.
- Availability.
- Shipping requirements.

1.3.3 Market Analysis for By-Products - Task 5.1.3

A study was performed with an initial objective of determining the potential sales volume and revenue for each potential by-product of the coal gasification facility.

By-products have been identified in Section 2.4, System Stream Characterization and Raw Materials. The existing market for each by-product has been identified and the potential demand for each by-product for these uses has been determined.

1.3.4 Alternate Products Analysis - Task 5.1.4

The objective of this task is to characterize and define the system/subsystems available to produce methane, methanol, hydrogen, or gasoline from an MBG feed stock and to characterize the associated cost components. Modifications to the integrated facility or additional equipment requirements, labor, and raw materials and their associated costs will be identified. The potential quantities and revenues of each alternate product previously identified will be developed as input suitable for life cycle costing and cash flow analyses.

Design studies and analyses already performed by the Team will be reviewed to summarize methods and costs for producing many of the by-products.

The alternate products/processes and costs will be summarized and presented with additional descriptive material and appropriate diagrams, flowsheets, schematics and references in a later report.

1.3.5 Preliminary Integrated Facility Requirements - Task 5.1.5

The objective of this task was to specify preliminary integrated facility requirements to serve as a basis for design of the reference facilities that will be used in the comparative evaluation of the TVA A/E designs.

All the previous subtasks will be compiled or collected into a preliminary narrative document which describes the Integrated Facility and the design/cost specifications required for each system/subsystem of the facility.

The facility requirements have been developed based on the above subtasks, TVA requirements, and the Team's previous design studies and experience.

The requirements for Texaco and K-T based plants have been summarized in two reports both titled "Preliminary Integrated Facility Requirements Document," in conformance with DR-9 specifications. Requirements include product specifications (gas pressure, temperature, composition, purity, volume per day), plant location, coal (proximate and ultimate analysis), by-product specifications, type(s) of gasifier and other specified process units, sparing philosophy, environmental and safety standards, raw material and product storage, plant performance (availability, load characteristics, turndown), purchased utility and water costs, and product gas cost.

This preliminary document provides a basis for providing (1) the Integrated Facility/Module definition; (2) the Integrated Facility Requirements Document; and (3) the Comparative A/E evaluation.

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2.0 SYSTEM CHARACTERIZATION

2.0 SYSTEM CHARACTERIZATION - TASK 5.1.1

The preparation of semi-definitive cost estimates and the performance of detailed process engineering calculations require that certain licensing or confidentiality agreements be in place with the owners of coal gasification and associated technologies in order to obtain information necessary to do this work.

In the absence of these agreements, studies and evaluations obtained in the public domain can be used as a basis for preparing preliminary process designs and budget level factored estimates. These budget level factored estimates are quite often used in the Hydrocarbon and Chemical Processing Industries to make decisions on further spending on or investigation of specific projects.

Usually, A/E firms performing studies on gasification projects for clients in the public domain, such as EPRI, DOE, Bureau of Mines, etc., have the licensing agreements to perform detailed engineering and cost estimates. The data presented in these study reports are then summaries of the engineering and cost estimates which provide a good scaling base when used with good engineering judgment.

2.1 Approach

The approach taken in preparing the system characterizations of this section has been to utilize the system designs and trade-off studies previously done by the BDM-Mittelhauser team as well as the other published and proprietary studies in the Team's libraries. These reference data have been reviewed, and those references which are most applicable to the reference facility have been identified and summarized in Appendix A, attached to this report.

2.2 System Identification

NASA-MARSHALL has provided definition of the candidate systems which comprise the integrated facility and have been characterized in this task. Table 2-1 identifies these Candidate Systems.

TABLE 2-1. NASA SYSTEM IDENTITY

1. COAL PREPARATION AND FEEDING
2. GASIFICATION - KOPPERS-TOTZEK COAL GASIFICATION PROCESS (TEXACO)
3. INITIAL GAS CLEANUP & COOLING
4. ACID GAS REMOVAL
5. SULFUR RECOVERY AND TAIL GAS TREATMENT
6. AIR SEPARATION
7. COMPRESSION
8. PROCESS SOLIDS TREATMENT (DEWATERING)
9. INCINERATOR
10. INSTRUMENTATION AND CONTROL
11. COAL HANDLING
12. SOLIDS WASTE RECYCLING/DISPOSAL
13. BY-PRODUCT PROCESSING
14. PLANT POWER SYSTEM
15. STEAM GENERATION/DISTRIBUTION
16. WATER SUPPLY
17. WATER COOLING SYSTEM
18. WASTE WATER TREATMENT
19. GENERAL FACILITIES
20. ALTERNATE PRODUCTS

TABLE 2-2. UNIT OPERATION CATEGORIES

NUMBER

10	COAL RECEIVING, STORAGE AND TRANSFER
11	COAL PREPARATION AND FEEDING
20	GASIFICATION
21	GAS COOLING
22	ACID GAS REMOVAL
23	COMPRESSION
31	SOLIDS TREATMENT SYSTEM
32	TAR-OIL SEPARATION
33	PROCESS CONDENSATE TREATMENT
34	PHENOL RECOVERY
35	AMMONIA RECOVERY
36	SULFUR RECOVERY
37	BIOLOGICAL TREATMENT
39	COOLING WATER SYSTEM
41	INCINERATION
80	AIR SEPARATION AND OXIDANT FEEDING
81	FINAL SOLIDS DISPOSAL
82	BY-PRODUCT STORAGE AND LOADING
83	SULFUR STORAGE AND LOADING
84	STEAM GENERATION
85	RAW WATER TREATMENT
86	FLUE GAS TREATMENT
87	PLANT ELECTRICAL SYSTEM
88	BUILDINGS AND SUPPORT FACILITIES
89	CONTROL AND INSTRUMENTATION

Data reported in the literature and studies has been based on a more discrete level of unit operations. In order to manipulate the reported data as little as possible, the BDM-Mittelhauser Team identified the pertinent unit operations that are typically included in a Coal Gasification Facility and obtained cost and system characterizations on that basis. Table 2-2 identifies the unit operations that were used to summarize data for this report. The summarized data has been reported on the unit operation level.

2.3 System Description

A brief description of each unit operation has been included in this section. Several unit operations have alternate processing technologies available for use in different gasification facility schemes. These alternates have been characterized to represent the choice of technologies available to the Integrated Facility which is yet to be designed.

Each description addresses the cost and design drivers of the unit operations, as well as the issues of critical technology addressed thus far in the study.

SYSTEM NUMBER 11
UNIT OPERATION NUMBER 10
DESCRIPTION COAL RECEIVING, STORAGE AND TRANSFER

The Coal Receiving, Storage and Transfer System provides for the unloading of coal delivered to the plant either by barge or truck, transporting the coal to storage, reclaiming the coal from storage, reducing the size of the coal, and transporting the coal to Coal Preparation, Unit Operation 11.

ROM Coal is unloaded from the barges by the barge unloading subsystem which is designed to unload up to ten 1500 ton capacity barges per shift. The coal is unloaded at an average rate of 1200 tons per hour on a 5 day/week basis. Coal is transferred by conveyor to a radial stacker which then forms a kidney shaped coal pile containing live and dead storage. Coal is reclaimed from the live storage and conveyed to a Bradford Breaker where it is reduced to 2"x0 size. Coal from trucks is unloaded into a chute from which it is conveyed to the Bradford Breaker. From there, the crushed coal goes to day storage silos from which it is transferred by vibrating belt conveyors to Coal Preparation, Unit Operation 11, for further processing.

Coal fines are collected and sent to Gasification (depending on process), Unit Operation 20; coal fired boiler for Steam Generation, Unit Operation 84; Solids Disposal, Unit Operation 81; or By-product Storage, Unit Operation 82, for sales.

The Design Drivers include:

- 1) Coal bulk density;
- 2) Blending requirements;
- 3) Angle of repose;
- 4) Equipment layout;
- 5) Sampling requirements;
- 6) Coal size distribution;
- 7) Barge or Rail delivery;
- 8) Coal moisture;
- 9) Coal hardness; and
- 10) Climate.

The Cost Drivers include:

- 1) Volume and shape of coal;
- 2) Available land area;
- 3) Conveyor length and width;
- 4) Construction materials;
- 5) Days of storage;
- 6) Concrete requirements; and
- 7) Equipment summary.

SYSTEM NUMBER 1
UNIT OPERATION NUMBER 11
DESCRIPTION COAL PREPARATION

This system receives raw coal from Coal Receiving, Storage and Transfer, Unit Operation 10; reduces the coal to the proper size; screens out and recycles the oversize fraction; and transfers the properly sized coal to Gasification, Unit Operation 20, for direct gasification or further treatment.

In some gasifiers, such as Lurgi, the Coal Preparation System is not part of the gasifier licensor's proprietary technology. In others, such as Texaco, final crushing and slurring is part of the proprietary package, while in the Koppers-Totzek process all coal crushing, drying and feeding is considered proprietary by the licensor.

Crushed coal 2"x0 is received by vibrating belt feeder from Coal Receiving, Storage and Transfer Unit Operation 10. Tramp iron separators then remove ferrous metal items from the coal stream and discharge them outside through the tramp iron chute. The coal is then weighed on a belt weigher after which it goes to a sizing crusher. The sized 1/4"x0 coal is stored in gasifier feed surge bins from which it is conveyed by vibrating belt feeders to Gasification, Unit Operation 20, at a total rate of about 20,000 TPD for the four modules.

Coal fines are collected and sent to Gasification (depending on process); coal fired boiler for Steam Generation, Unit Operation 84; Solids Disposal, Unit Operation 81; or By-product Storage, Unit Operation 82, for sales.

The Design Drivers include:

- 1) Fines minimization (Lurgi) - product coal size;
- 2) Coal source (ROM or washed);
- 3) Blending requirements;
- 4) Hardgrove grindability;
- 5) Angle of repose;
- 6) Moisture content;
- 7) Coal bulk density;
- 8) Sampling requirements; and
- 9) Waste heat availability for coal drying.

The Cost Drivers include:

- 1) Gasifier feed coal size requirements
- 2) Plant layout
- 3) Coal source (ROM or washed).

Critical technologies identified as needing study and development work are:

- Wet coal grinding - wet grinding applies to the Texaco gasification system. Maximization of pumpable coal slurry concentrations leads to improved thermal efficiency. Needs include identification of optimum particle size distributions, methods of producing optimum distributions, and the effect of coal blending.
- Coal Blending - Needs in coal blending include automatic blend control technology, data on effect of blending on gasifier slag properties, and effect of blending on gasifier residence time requirements.
- Recycle of waste water - Utilization of waste water for coal slurry preparation is one method of utilizing this material. Needs include identification of the buildup of toxic materials in recycled waters and methods of controlling same through side-stream treatment or blowdown.

SYSTEM NUMBER 2
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - KOPPERS-TOTZEK

The Koppers-Totzek gasifier is a high temperature, cocurrent entrained flow gasifier which accepts coal from Coal Preparation along with oxygen and steam to produce intermediate BTU gas. It is a proprietary unit licensed by Krupp-Koppers of Germany. Sized coal 1/4"x0 from Coal Preparation, Unit Operation 11, enters the pretreatment area of Gasification, Unit Operation 20, where it is crushed and ground to 70% minus 200 mesh, and dried to 2% moisture. It is then fed to eight screw conveyors that feed four pairs of burners located 90° apart. There are four feed points on the four-headed gasifier with 2 burner heads at each point. Each burner projects a jet to converge with the line of discharge of the other. Oxygen from Air Separation, Unit Operation 80, and steam from Steam Generation, Unit Operation 84, carry the coal through the burners into the gasifier.

The oxygen, steam and coal react to gasify the carbon and volatile matter of the coal and to convert the coal ash into molten slag which is sent to Solids Treatment, Unit Operation 31. The gas exiting each gasifier is direct water quenched to below the ash fusion temperature, in order to solidify entrained slag droplets. The remaining slag forms a layer on the refractory walls and flows down through a separate chute into quench tanks.

Low pressure steam is produced in the water jackets of the gasifiers from waste heat that passes through the refractories.

After the gas is quenched, gas and entrained ash particles pass through a waste heat boiler where the gas is cooled to approximately 350°F by raising high pressure steam. The gas is then scrubbed for particulate removal. The clean intermediate BTU product gas is then further cooled in Gas Cooling, Unit Operation 21, before going to Acid Gas Removal, Unit Operation 22.

With the K-T gasifier, as with all high temperature entrained flow gasifiers, no tars, phenols, oils, etc., are produced so the gas requires less cleanup than those systems that produce hydrocarbons. Because of the high operating temperatures the gasifier requires an appreciable amount of oxygen per pound of coal fed. The K-T gasifier requires about 0.9-1.0 lb O₂ per lb coal fed. Steam consumption is approximately 0.4 lb per lb of coal. The higher heating value of the dry gas produced from the K-T gasifier is in the range of 285-300 BTU/SCF. The Koppers-Totzek gasifier typically operates at a pressure of about 7 psig. Maximum temperatures can run as high as 3300°F.

The Design Drivers for the system include:

- 1) Unit capacity, TPD coal
- 2) Coal (especially ash) properties.

The Cost Drivers for the system are the same as the Design Drivers.

SYSTEM NUMBER 2
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - TEXACO

The Texaco Coal Gasification Process uses a coal slurry feed, consisting of fresh ground coal together with recycled fine slag and carbon with a total solids content 50 to 65% by weight. The slurry is pumped from mix tanks in the grinding and slurry section to the gasifier slurry tank. A circulating pump circulates the slurry through this tank and supplies slurry to the suction of the high pressure charge pump.

The coal-water slurry is fed through a specially developed burner into a refractory-lined gasifier reactor. Partial combustion with oxygen takes place at a pressure of 600 psig, or higher, and a temperature in the range of 2300 to 2800 degrees F to produce a gas consisting mainly of CO, H₂, CO₂, and steam. Most of the sulfur in the coal is converted to H₂S and the balance converts to COS. Nitrogen and argon from the oxygen feed appear in the gas together with most of the nitrogen from the coal. The gas contains a small amount of methane, some unconverted carbon and all of the ash in the form of slag. The gas is essentially free of uncombined oxygen.

The upper section of the gasifier is the refractory-lined chamber in which the partial oxidation reaction takes place. In many conceptual designs, part of the gas is withdrawn and cooled to below the ash fusion point by mixing with cooled recycle gas. Entrained slag particles, solidified by cooling, are then removed from the gas. The gas is then cooled by raising high-pressure steam in a specially-designed waste heat boiler. The gas then passes to the Gas Cooling System, Unit Operation No. 21. To date, these high-pressure steam generators have not been commercially proven in coal gasification service.

At least a portion of the gas from the gas generator reaction section passes straight down into the quench section of the gasifier. This stream carries the bulk of the larger particles of slag, and it is immediately quenched with water from the 2300 to 2800 degree F range to about 400 degrees F. The gas from the generator quench chamber joins the main stream of gas going to the gas cooling operation.

Water from the gasifier quench chamber is cooled and combined with water from the carbon-scrubber lower section. Both streams contain fine slag and unconverted coal. The water stream from the lock hopper and water from the final product cooler separator join this stream, and the total flows into the flash pot.

In the clarifier, the fine slag and unconverted coal settle out, leaving a clarified water overflow that is pumped back to the carbon scrubber via the gray water drum. Makeup water is added at this point.

The clarifier underflow is fed to a centrifuge for dewatering. The concentrated underflow is returned to the coal grinding and slurring section and mixed with coal-feed slurry. The filtrate is returned to the clarifier.

Most of the ash in the coal feed agglomerates into essentially carbon-free molten slag droplets, which are quenched and solidified in the lower quench section of the reactor. This slag is settled through the quench water into the lock hopper. The lock hopper is periodically dumped onto a screen, from which the slag is conveyed to the solids treatment system.

The primary cost drivers are:

- Unit capacity, TPD coal
- Coal (especially ash) properties
- Feed slurry concentration
- Operating pressure.

The major design drivers are the same as the cost drivers.

SYSTEM NUMBER 2
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - LURGI

The Lurgi gasifier, dry ash, gravitating bed type, is commercially available from Lurgi Kohle and Mineraloeltechnik. The gasifier is a water jacketed pressurized unit comprised of a series of vertically stacked vessels. There are, from top to bottom, a coal hopper, coal lock, water jacketed gasifier, ash lock and ash quench chamber.

Coal is conveyed from Coal Preparation, Unit Operation 11, to the coal hopper from which it is fed by gravity to the depressurized coal lock through a hydraulically operated valve. The lock is then isolated and pressurized with a slipstream of inert gas (mainly N_2) and the coal is transferred to the gasifier through another hydraulically operated valve. The empty lock is isolated, depressurized through a bag filter and vented either to the atmosphere or the Incinerator, Unit Operation 41. The gas displaced from the coal and lock hoppers during loading is routed similarly. Coal dust recovered in the filter is returned to the coal hopper.

The coal flowing down through the gas produced represents a slowly moving bed which has several distinct zones. In the first zone at the top of the gasifier, coal is preheated and dried by contact with the hot crude gas leaving the reactor. As the coal moves down and is heated further, devolatilization occurs and gasification commences. The bottom of the bed is a combustion zone where carbon reacts with oxygen to form CO and CO_2 . The oxidation provides the overall heat for the gasification and devolatilization reactions which are endothermic. Only a negligible amount of unburned carbon remains in the ash.

When MBG is to be made, oxygen from Air Separation and Oxidant Feeding, Unit Operation 80, and steam enter the gasifier near the bottom and are heated as they rise upward to the combustion zone by the hot ash moving down from the combustion zone. Oxygen flow rate is controlled to accomplish complete gasification of coal. Steam rate is controlled to maintain a specified gasifier bottom temperature to prevent melting or clinkering of the ash.

A portion of the gasifier process steam is generated at about the operating process of the gasifier, in the gasifier jacket. The balance is provided through waste heat recovery or from Steam Generation, Unit Operation 84.

The crude gas leaving the gasifier contains appreciable quantities of tars, oils, naptha, phenols, fatty acids, ammonia, hydrogen sulfide, sulfur compounds and a small amount of coal and ash dust. The crude gasifier effluent temperature ranges from 575°F to over 1000°F. The effluent flows through a scrubbing cooler where it is washed with a stream of process condensate. The washing process quenches the gas to about 350 - 400°F and condenses the high boiling tar fractions. Coal and ash dust are removed

with the condensed tar leaving the quenched effluent gas essentially free of particulate matter.

Ash from the process is continuously collected by a rotating ash grate and moved to the ash lock hopper. Ash collected in the lock is depressurized and discharged batchwise to an ash quench chamber where it is cooled in water. The ash lock is pressurized with steam.

The abrasive slurry from each gasifier train flows to a common transfer tank using water as the motive fluid. Ash grinders are provided to prevent large chunks of slag from plugging transfer lines. The ash slurry is then sent to Solids Treatment, Unit Operation 31.

The primary cost and design drivers are:

- Coal and Ash properties
- Capacity, TPD of coal
- System pressure

SYSTEM NUMBER 2

UNIT OPERATION NUMBER 20

DESCRIPTION GASIFICATION - BABCOCK AND WILCOX

The Babcock and Wilcox gasifier is a high temperature, cocurrent entrained flow gasifier which accepts coal from Coal Preparation along with oxygen and steam to produce medium BTU gas. It is a proprietary unit licensed by Babcock & Wilcox.

Sized coal 1/4"x0 from Coal Preparation Unit Operation 11, enters the pretreatment area of Gasification, Unit Operation 20, where it is pulverized to 70% minus 200 mesh and tangentially injected through two rows of water cooled nozzles into the gasifier. Recycled char from the gasifier outlet gas cyclones is also injected through water cooled nozzles in the bottom row of burners. Both the coal and char are fired with oxygen from Air Separation, Unit Operation 80. The coal and char are partially combusted to form a hot reducing gas. At the high temperatures present in the gasifier, the ash in the coal and char becomes molten and continuously flows down the walls of the gasifier to the slag tap hole. From the tap hole, the slag enters a water quench tank where it is cooled. From the quench tank it flows to the Solids Treatment System, Unit Operation 31.

In the gasification section, there is an inner shell of water cooled tubes (water wall) where saturated steam is produced. In the hot reaction zone, the tubes are covered with a dense refractory suitable for contact with molten flowing slag. Above the reaction zone, the tubes are bare for greater radiation cooling prior to entrance into the waste heat boiler section.

The gas exits the gasifier proper at about 1800°F and enters the waste heat boiler section where it is cooled to 700°F. From the waste heat boilers, the gas enters a cyclone where 90-95% of the carryover ash and char is removed. This char and ash stream, as mentioned previously, is injected back into the gasifier. The 700°F gas is further cooled and cleaned in Gas Cooling, Unit Operation 21, before going to Acid Gas Removal, Unit Operation 22.

The Design and Cost Drivers include:

- 1) Coal (especially ash) properties
- 2) Capacity, TPD coal
- 3) Operating pressure.

The Cost Drivers include:

- 1) Gasifier pressure; and
- 2) Ash fusion temperature.

SYSTEM NUMBER 2
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - SLAGGING LURGI

This system is divided into the following four subsystems:

- Coal and Flux Feed
- Gasification
- Raw Gas Treating
- Slag Handling.

Coal and Flux Feed

Flux is an agent which forms a eutectic mixture with the coal ash in the gasifier, lowering its melting point to make slag formation easier. It is shipped to the facility from outside sources.

The coal and flux are mixed in Coal Preparation Unit Operation 11 and then fed to the coal bunkers by a belt conveyer system. The feed chutes at the bottom of the coal bunkers control the flow of coal into the coal locks. Each gasifier has two coal locks that operate automatically on a cyclic basis. There, coal locks are pressurized with mostly N_2 and alternately feed the coal surge vessel.

Gasification

The design of the gasifier is based on proprietary technology held by Lurgi Kohle and Mineraloeltechnik and the British Gas Corporation. It is similar to the dry-ash Lurgi gasifier described earlier, except that in the bottom of the gasifier the coal ash melts as a eutectic with the added flux to form slag. The molten slag collects at the bottom and is removed intermittently from the gasifier through a slag tap hole.

The coal and flux, entering the top of the Gasifier, descends in a moving bed in countercurrent flow to steam, oxygen and produced gas. While traveling from the top to the bottom of the Gasifier, the coal is dried, devolatilized, and gasified. The heat required for these three steps is supplied by the exothermic reaction between the carbon in the coal and the oxygen in the bottom of the Gasifier.

As the produced gas passes through the coal bed, its final composition is determined by the following:

- Exothermic and endothermic reactions occurring simultaneously in the gasification zone,
- Formation of hydrocarbons, phenols, fatty acids, and minor organic compounds in the devolatilization zone, and
- Evaporation of coal moisture in the drying zone.

Raw Gas Treating

Raw gas is treated similarly to that from a dry-ash Lurgi gasifier, as described earlier.

Slag System

After the coal ash melts as a eutectic with the added flux to form slag, the molten slag collects at the bottom of the gasifier and is tapped intermittently through a tap hole into the Quench Vessel. In the Quench Vessel, the slag granulates immediately upon contact with the quench water. The granulated slag falls into the Slag Hopper and is dumped once or twice an hour. The slag and water mixture that is dumped goes to Solids Treatment Unit Operation 31.

The Design Drivers include

- 1) Coal and Ash Properties
- 2) Gasifier Pressure.

The Cost Drivers include

- 1) Coal and Ash Properties
- 2) Gasifier Pressure
- 3) Capacity, TPD coal.

CRITICAL TECHNOLOGY - UNIT OPERATION 20

Critical technologies identified as needing study and development work are:

- Gasifier Instrumentation - Needs in gasifier instrumentation include temperature monitoring throughout the gasifier, early detection of reactor hot spots, quick response control of reactant flow controls. Reliability of reactor instrumentation is a major safety factor in entrained gasifiers.
- Reactor effluent fines removal - Efficient and reliable fines removal from gaseous products is required for protection of downstream units. Needs include materials of construction.
- Reactor Refractory - Development of improved reactor refractory is needed to increase time between repairs. Needs include data on which coal properties or constituents influence the refractory life, and development of improved materials, application methods, or configurations.
- Slag Handling - Needs include slag flow control techniques for slag draining. Slag viscosity control is also desirable especially in case of gasification of coal blends.
- Instrumentation - The appropriate degree of automatic closed-loop control versus human set-point control needs to be tested and established for different types of gasifiers and load situations.
- Product Data - Improved methods for predicting nitrogen and sulfur product spectrums from coal properties and gasification process conditions would be a substantial aid to designing downstream equipment.

SYSTEM NUMBER 3

UNIT OPERATION NUMBER 21

DESCRIPTION

RAW GAS COOLING

The purpose of this unit is to cool the gasifier effluent gas to more amenable processing temperatures in subsequent systems and to separate ash and solids from the gas. This system description is dependent upon gasifier selection, as some gasifiers include one or more stages of cooling as an integral part of the gasifier system.

The portions of gas cooling associated with the gasifier are proprietary and are handled by the gasifier licensor. The remaining portions of gas cooling are non-proprietary.

Gas cooling systems generally consist of heat exchangers and pressure vessels. The removal of heat from a gas mixture containing tars, oils, and dust requires special design considerations to prevent plugging of tubes and excessive fouling of cooling surfaces. These include vertical tubes in exchangers with process gas on the tube side, and washing of the tube walls where required with reinjected gas liquor.

The pressure vessels are either simple vapor-liquid separators or quench vessels in which water is sprayed countercurrent to the gas to cool it and to remove solids, oils, tars and some of the water vapor from the gas.

The primary cost drivers are:

- Gas quantity, and
- Acid gas removal system temperature requirements.

The primary design drivers are:

- System pressure
- Heat integration requirements.

Critical technology areas needing study and development work are:

- High temperature heat recovery - Recovery of heat at higher temperatures will make possible designs of more efficient plants. Needs include heat transfer data on both water and gas sides of exchangers.
- Long term reliability of construction materials.
- Distribution coefficients for gasifier products - Additional data on gaseous constituents and trace metal compounds are needed for efficient design of downstream equipment.

SYSTEM NUMBER 4
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - SELEXOL

The purpose of the Selexol unit is to remove H_2S , other sulfur compounds, and CO_2 from the raw gas by physical absorption. This is a proprietary process developed and licensed by Allied Chemical Corporation.

The Selexol process removes sulfur compounds and carbon dioxide by counter-current contact in a packed absorber using dimethyl ether of polyethylene glycol as the solvent. The treated gas from the top of the absorber flows through a knockout drum, and then leaves the system.

The rich solvent from the bottom of the absorber is usually let down in pressure through a hydraulic turbine, which supplies a portion of the power required by the lean solvent solution pump. It then flows to one or more flash drums where most of the dissolved hydrocarbon gases in the solvent flash off. Most of the dissolved H_2S and COS are retained in the solvent because of their selective absorption in the Selexol solvent. Flashed gas is compressed and recycled to the gas feed to the absorber.

The rich solvent solution from the flash drum is preheated by hot regenerated solution in a heat exchanger and flows to the top of a stripper, where absorbed H_2S , COS , and CO_2 are stripped from the solution. Reboil heat is supplied by low pressure steam in a shell-and-tube reboiler. Hot lean solvent first exchanges heat with rich solvent in order to reduce reboiler duty and is then pumped back to the absorber. The lean solution is cooled down to operating temperature with cooling water in a shell and tube exchanger.

Acid gas (H_2S , COS , and CO_2) from the stripper overhead is cooled to $120^\circ F$ in either an air fan cooler or against cooling water. The condensate produced in cooling is separated in a knockout drum, and then pumped back to the stripper. The cooled acid gas flows to sulfur recovery, Unit Operation 36, for further processing.

The Design Drivers include:

- Gas composition
- Treated gas purity specifications
- Sulfur recovery unit H_2S content requirements
- Absorption pressure.

The Cost Drivers include:

- Inlet Gas Capacity, ACFM
- Moles/hr H_2S , COS , CO_2 removed
- Selective (double acid gas stream) or nonselective (single acid gas stream) removal
- Absorption pressure.

SYSTEM NUMBER 4
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - STRETFORD

The Stretford Process is used for removal of Hydrogen Sulfide (H_2S) at essentially atmospheric pressure. This process is a direct oxidation process which absorbs H_2S from the crude gas in an alkaline solution of sodium metavanadate and anthraquinone disulfonic acid and converts it to elemental sulfur. H_2S can be removed to less than 1 PPM in the treated gas.

The Stretford Process is a proprietary process licensed by North West Gas Board, Ltd., and offered by Parsons and others in the U.S.

The quenched raw gas from Gas Cooling, Unit Operation 21, is fed to either a venturi scrubber followed by a small packed absorber or into a single larger packed absorber. The hydrogen sulfide is absorbed by sodium carbonate to form sodium hydrosulfide.

An inlet temperature of below 120°F is required to minimize the formation of the unreactive sodium thiosulfate which must be purged from the solution. The "rich" solution from the absorption step goes to a reaction tank where sodium metavanadate oxidizes the hydrosulfide to elemental sulfur.

The four-valent state of the vanadium reverts back to the five-valent state due to the presence of anthraquinone disulfonic acid (ADA).

The mixture of sulfur and Stretford solution next flows to the oxidizer. In the oxidizer, the ADA is reoxidized with air and the sulfur is floated and skimmed from the solution. The sodium hydroxide and bicarbonate produced by the above sections combine to form sodium carbonate. The regenerated solution is returned to the absorption step.

The skimmed sulfur is sent to a sulfur purification system. Here, the sulfur float containing around 10 weight percent sulfur is further processed in filters or centrifuges to separate the solution from the froth. The filter cake, containing 50 to 60 percent solids, can be further processed by melting in an autoclave to produce high grade sulfur which is sent to sulfur by-product storage, Unit Operation 83.

The Stretford process can reduce the H_2S content of treated gas to less than 50 ppm, but it does not absorb CO_2 .² Also, it has not yet been commercially proved on coal-derived gases at pressures much above atmospheric. The Stretford process has been proven in many installations including coke oven gas treatment. Potential problems associated with its use in gasification are the presence of HCN, organics, and particulates in the gas. The HCN reacts in solution to form NaSCN which is stable and must be purged from the solution, resulting in high solvent makeup costs. If the purge increases to the point that the Stretford process is not cost effective,

then the HCN must be removed by costly polysulfide wash or catalytic oxidation steps. If organics and particulates are present in the quenched gas, they could be removed by the Stretford scrubber. If this occurs, these materials could discolor the sulfur and impair its marketability.

The primary design drivers in the Stretford process are

- 1) Mole percent CO_2 in the feed gas
- 2) Absorption-temperature
- 3) Feed gas HCN, organics, particulates
- 4) Feed gas COS content.

CO_2 is partially absorbed by the alkaline solution which reacts to form bicarbonates and lower the pH of the solution. When feed gases contain high concentrations of CO_2 , the absorption efficiency of the solution may be sufficiently lowered to require an appreciable increase in packed absorber height. The venturi/absorber system can remedy this problem.

The primary cost drivers in the Stretford Process are

- 1) Moles/hr sulfur equivalent removed
- 2) Mole % HCN in the feed gas.

SYSTEM NUMBER 4
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - BENFIELD

The purpose of the Benfield unit is to remove H_2S and other sulfur compounds, and CO_2 , from the raw gas by means of chemical absorption. This is a proprietary process developed and licensed by Benfield Corporation, a subsidiary of Union Carbide Corporation.

The Benfield process removes sulfur compounds and carbon dioxide by countercurrent contact in a packed absorber using an aqueous solution of potassium carbonate at a temperature of 200-230°F, by a proprietary additive. This additive promotes the chemical reaction between the solvent and the H_2S , COS , and CO_2 to be absorbed. A rich solvent from the absorber passes directly into a stripper where the sulfur compounds are stripped from the solution by reboiling against low pressure steam.

Because the H_2S , COS , and CO_2 are chemically combined with the solvent, pressure reduction is not as effective in removing these compounds from the solvent as is the case with a physical solvent.

After stripping, the lean solvent is cooled slightly and returned to the absorption step.

To achieve deep removal of H_2S and CO_2 , a proprietary modification known as the "Hi-Pure" Process is used. This process uses two independent but compatible circulating solutions in series to obtain both high purity treated gas and high thermal efficiency. The gas is first contacted with normal hot potassium carbonate to remove the bulk of the acid gases. Final purification is then achieved in a separate absorber using a solution of somewhat different composition. The two solutions are regenerated separately in two sections of a regenerator with the stripping steam leaving the lower section of the regenerator being re-used in the upper section.

Primary design drivers for the Benfield Process are:

- Gas Composition, especially H_2S , COS , CO_2
- Treated Gas Purity Specification
- CO partial pressure in feed
- Absorption Pressure.

Cost Drivers include:

- Inlet Gas Capacity, ACFM
- Moles/hr acid gas removed
- Absorption Pressure.

SYSTEM NUMBER 4
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - RECTISOL

The purpose of the Rectisol System is to purify or adjust the composition of the raw gas for subsequent processing.

This is a proprietary technology licensed by Lurgi Kohle and Mineraloel-technik, GmbH.

The Rectisol process removes H_2S , COS, CO_2 , and naphthas (in gas from Lurgi gasifiers) using refrigerated methanol as a physical solvent. The system may be comprised of the following subsystems:

- Gas Cooling
- Naptha Recovery (optional)
- Methanol Recovery
- Raw Gas Treating
- Flash Regeneration
- H_2S Concentration
- Hot Regeneration
- CO_2 Recovery/Compression
- Refrigeration.

Gas Cooling

This section is used to precool the gas from Gas Cooling, Unit Operation 21, by heat exchange against refrigerant and cold processed gas. The temperature of the gas is reduced from about 100-120°F to well below ambient temperature.

Naptha and Methanol Recovery

The Naptha recovery section, located upstream of the raw gas treating section, can be used to remove organic sulfur (COS) and naphtha-range hydrocarbons from the cooled gas. This is done with a small quantity of cold methanol which is then partly regenerated by pressure reduction. Naptha is separated from the methanol by extracting the methanol in a wash water stream. The aqueous stream is regenerated by steam stripping and the methanol recovered in the Methanol Recovery Section by distillation, for recirculation to the absorption steps.

Raw Gas Treating

In this step, H_2S , COS, and CO_2 are absorbed by refrigerated methanol. A complex sequence of absorption, flash, and heat exchange equipment is used to regenerate the solvent and produce lean methanol for recirculation to the absorption steps.

Flash Regeneration

Flash regeneration is a complex sequence of pressure reduction steps by which absorbed light gases (H_2 , CO, CH_4) are recovered and compressed for recycle to the main MBG stream.

H_2S Concentration

This section flashes off CO_2 from the rich methanol so that the H_2S concentration in the acid gas going to Sulfur Recovery Unit Operation 36, will be maximized, to around 50 volume percent.

Hot Regeneration

This section strips the remaining H_2S , COS, CO_2 and hydrocarbons from the methanol by steam stripping. The Acid Gas overhead is cooled to recover methanol and then goes to Sulfur Recovery, Unit Operation 36.

CO_2 Recovery/Compression

CO_2 flashed from the rich methanol in the H_2S concentration section is compressed in the CO_2 Recovery/Compression Section for use as lock gas in Gasification, Unit Operation 20. If not needed for this purpose, the gas may be vented.

Refrigeration

This section uses liquid propylene in a vapor compression cycle to provide subambient cooling required for the remainder of the system.

The Design Drivers are:

- 1) H_2S /COS/ CO_2 Feed
- 2) H_2S /COS/ CO_2 Removal required
- 3) Sulfur Recovery Method
- 4) Organics/Trace Components in Feed.

The Cost Drivers are:

- 1) Inlet Gas Capacity, ACFM
- 2) Absorption Pressure
- 3) H_2S /COS Removal moles/hr
- 4) Solvent Loss
- 5) Selectivity requirements.

SYSTEM NUMBER 7
UNIT OPERATION NUMBER 23
DESCRIPTION

COMPRESSION

The purpose of this unit is to compress and dry the MBG gas produced for delivery to the pipeline at 600 psig. The dry compressed gas is usually metered before delivery to the pipeline.

The compressors may be located at various positions within the overall processing sequence. Regardless of where the gas is compressed, the drying unit must be the last unit in the IBG processing sequence.

The condensate from the compressor suction interstage and discharge drums is returned to the steam generation plant for use as boiler feed water. Water vapor from the gas drying unit is vented to the atmosphere.

The compressed gas is dried to meet a specified maximum water content of seven lb/MMSCF gas by a conventional triethylene glycol (TEG) system. The TEG system is designed to normally achieve a water level of six to seven lb/MMSCF pipeline gas.

This is a non-proprietary process that can be designed/supplied by several US manufacturers.

The Design and Cost Drivers are:

- 1) Gas Properties
- 2) Pressure Boost and Suction pressure
- 3) Gas Flow Rate, SCF per hour.

Critical technology areas needing further study and development work are:

- Shaft seals for hydrogen systems - Good sealing of shafts against hydrogen leakage is difficult. Development work is needed for seal systems in large compressors.

Critical technology areas needing further study and development work are:

- COS removal - Removal of COS is required both for protection of downstream catalytic systems and product specification requirements. Satisfactory techniques are not now well developed.
- Selective Solvents - Development of more selective solvents would increase process efficiency and decrease capital costs.
- Solvent Degradation - Little is known about solvent degradation caused by minor constituents in coal gasification products.
- Distribution of Organics - Little testing has been done concerning the distribution of organics in coal gasification acid gas removal systems.

SYSTEM NUMBER 8
UNIT OPERATION NUMBER 31
DESCRIPTION SOLIDS TREATMENT SYSTEM

The purpose of this unit is to collect and dewater the various solids slurries, or sludges resultant from the Facility Operation for economical, environmentally acceptable disposal.

The processes involved in this system are nonproprietary and are supplied by various US vendors.

Typically, ash and slag from the gasifier and gas cooling system, biological sludges, and solid wastes from process condensate treatment are treated in this unit. The treatment methods below can be used to accommodate the solids treatment.

Gravity settlers - used to separate dense solids from the waste streams.

Float thickness-clarifiers - used to treat slurry from the gravity settlers. Thickeners and coagulant aids are added to facilitate solid-liquid separation.

Rotary Drum or Belt Filters - filters sludges from the process condensate treating systems or from the float thickness.

Recovered water is sent to the Process Condensate Treating System, Unit Operation 33, and solids are conveyed to Final Solids Disposal, Unit Operation 81.

The Design Drivers are:

- 1) Sludge Feed Characteristics.

The Cost Drivers are:

- 1) Capacity
- 2) Type Dewatering Equipment
- 3) Transport Distance
- 4) Disposal Method
- 5) Storage Requirement.

Critical technology areas needing further study and development work are:

- Trace metals in cooling/quench water - Direct contact between slag and cooling/quench water provides an opportunity for trace metal contamination of waste water. Data on concentrations is needed. Methods of removal and/or recovery of trace metals need developing.

- Leachability of metals - Data on leachability of metals from ash/slag disposal pits is needed.
- Zero discharge - Zero discharge concept needs demonstrating. Process for concentration of waste contaminants by vapor compression and evaporation need developing.

SYSTEM NUMBER 13

UNIT OPERATION NUMBER 32

DESCRIPTION

TAR-OIL SEPARATION

This unit is used in the Lurgi and BGC-Slagging Lurgi Processes to collect the condensate from Gasification, Shift Conversion, Gas Cooling, and Rectisol. The condensate, called gas liquor, contains tar, oil, dust and other impurities. Gravity settling tanks are used to separate the tar, oil and dust from the gas liquor. The tar and dust are recycled to Gasification so there is no net production. The oil is recovered either as a saleable product or for plant fuel.

This design is based on proprietary technology held by Lurgi Kohle and Mineraloeltechnik, GmbH.

Dusty and oily gas liquors from the waste heat exchangers in Gasification and Shift Conversion are combined and cooled in an air cooler before being expanded to atmospheric pressure. The pressure drop causes gas to be released from the liquor. The gas is separated from the liquid in the Dusty Gas Liquor Expansion Drums. The gas liquor flows to the Primary Gas Liquor/Oil/Tar Separators where the dusty tar is withdrawn from the bottom. Clear tar flows from the top of the separators over an adjustable overflow to the Tar Tanks. The oil flows with the gas liquor to the Secondary Gas Liquor/Tar/Oil Separators by gravity where more tar is removed and drained to the Tar-Oil Slop Tank. Oil is removed from the secondary tar separators via the adjustable overflow to the Tar Oil Tank. The gas liquor is used for injection water in gasification.

Oily gas liquor from Gas Cooling and gas liquor from Rectisol are cooled prior to entering the Oily Gas Liquor Expansion Vessels. The gaseous components are released in the expansion vessels and the gas liquor flows to the Oil Separators. The oil is withdrawn from the separators over an adjustable overflow.

The dusty tar withdrawn from the bottom of the Primary Gas Liquor/Tar/Oil Separators is pumped into the top of the gasifiers.

Clear tar from the Tar Tanks is pumped in a similar recycle by the Clear Tar Pump to the Gasification unit.

The primary cost and design driver is the quantity of tar and oil removed from the effluent gas.

Critical technology areas needing further study and development work include:

- Tar and oil properties - Correlation between tar/oil properties and coal type are not available. Development of such correlations as well as correlations with gasifier conditions would

improve engineering design and/or reduce the required lab work preceeding a detailed design.

- Distribution coefficients - Distribution coefficients are needed for all phase separation steps in the tar/oil recovery systems especially aqueous constituents.
- Physical properties - Better data on physical properties of these systems, such as viscosity and surface tension, are needed.

SYSTEM NUMBER 18
UNIT OPERATION NUMBER 33
DESCRIPTION PROCESS CONDENSATE TREATMENT

The purpose of this unit is to collect and treat all facility liquid effluent streams. The facility design is predicated on "zero discharge." A wastewater treatment system permitting recycle and reuse of treated water would be required to meet this requirement.

This system contains various non-proprietary processes that can be supplied by several US vendors.

The number, type, quantity, and composition of liquid effluents generated within the facility are dependent on the gasifier technology selected and the process systems selected to produce the end product(s). The following list identifies the possible liquid effluents to be treated and, in some cases, may be mutually exclusive:

- Oily Water Sewers
- Coal Pile Run Off
- Storm Water Run Off
- Demineralizer Regenerant Wastes and Rinse Water
- Cooling Tower Flowdown
- Boiler (Steam Generator) Blowdown
- Ammonia Recovery System Blowdown
- Rectisol Blowdown
- Sanitary Waste Water
- Flue Gas Treatment Slurry
- Gasifier Slag Quench Drains
- Separated Water From Solids Treatment
- Filtrate From Biological Treatment.

Process operations (subsystems) would be selected to treat particular waste streams identified from the above list, depending on facility design. The following subsystems are typically selected or designed to treat the waste streams.

OIL SEPARATOR - streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemicals and heat to separate the oil-water mixture.

SCOUR WATER STRIPPER - water streams with appreciable H_2S or NH_3 residuals are steam stripped to remove these contaminants.

EQUALIZATION BASIN - liquid streams with extremely high or low pH are mixed in an equalizing basin and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0.

GRAVITY SETTLING-THICKENER - liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening.

MULTIPLE EFFECT EVAPORATION - neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make up to cooling towers or raw water supply. The resultant solids are conveyed to the Solids Disposal System.

The primary cost and design drivers are number, quantity, and types of liquid effluents to be treated.

Critical technology areas needing further study or development are:

- Process calculations - Both data acquisition and calculation methods development are needed for aqueous systems containing electrolytes and organic constituents.
- Equipment construction - Less expensive materials of construction need developing. The rigorous process design of water strippers is not now well developed.

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 34
DESCRIPTION PHENOL RECOVERY - PHENOSOLVAN

The Phenosolvan process is used to treat phenol-containing water from Process Condensate Treatment, Unit Operation 33. The feed stream has a temperature of about 150 degrees F and a pressure around 15 PSIG. Phenol is recovered for sales, and dephenolized water is transferred to Ammonia Recovery, Unit Operation 35, for further processing.

The Phenosolvan process is a proprietary process licensed by and offered by American Lurgi Corporation (N.Y.).

The process recovers phenols from the feed stream in a series of mixer-settler tanks. Within the mixer-settler the feed is contacted in a countercurrent liquid-liquid extraction operation, with isopropyl ether acting as the solvent. Phenols are extracted into the solvent with only 10-20 ppm (wt) of phenols remaining in the product water effluent.

The extract (containing the phenols along with solvent) is sent to a distillation column where the solvent is recovered and recycled. The phenol (80-90% pure) is transferred to Byproduct Storage, Unit Operation 82, for sales.

The product water from the mixer-settler is saturated with solvent. The solvent is recovered in a stripper and recycled. The phenol and solvent free water is sent to Ammonia Recovery, Unit Operation 35.

The primary design drivers in the Phenosolvan process are:

- 1) Feed water flow rate, temperature, and pressure; and
- 2) Solvent recovery efficiency.

The primary cost drivers are:

- 1) Percent phenol recovered
- 2) Mole percent phenol in feed.

Critical technology areas needing further study and development are:

- System performance data - Phenol recovery systems are now primarily associated with the steel industry. Performance data are needed on coal gasification waste water treatment systems.

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 35
DESCRIPTION . AMMONIA RECOVERY - PHOSAM=W PROCESS

This is a proprietary process licensed by USS Engineers, Inc.

This unit receives sour water from Phenol Recovery, Unit Operation 34 (if present), containing volatile components such as ammonia, carbon dioxide, hydrogen cyanide, and hydrogen sulfide. The feed enters a stripper where the volatile components are stripped by reboiling with indirect low pressure steam. This overhead stream is cooled and the condensate returned as reflux to the stripper.

The stripper bottoms being essentially ammonia-free is cooled by preheating the incoming feed, further cooled, and then sent to Biological Treatment, Unit Operation 37, if residual organics are high enough to warrant such treatment. The overhead vapor or "rich gas" is sent to an absorber where the ammonia is selectively absorbed in a circulating aqueous proprietary solution. The overhead gases are partially condensed. The condensate, which will contain some ammonia, is sent back to the stripper as part of the reflux. The non-condensed vapors containing a large fraction of carbon dioxide, hydrogen cyanide, and hydrogen sulfide are sent to Sulfur Recovery, Unit Operation 36. The ammonia-rich solution is sent to the top of the ammonia stripper and is partially heated by cross-exchange with the recycled lean solution. The ammonia is reboiled out of the solution which leaves the bottom of the ammonia stripper. This stream is cross-exchanged with the rich solution, further cooled, and recycled back to the top of the absorber.

The wet ammonia vapors from the top of the stripper are condensed, and the condensate is pumped into a fractionator. The overhead vapors are condensed, and the condensate (anhydrous ammonia) is sent to By-Product Storage, Unit Operation 82, for sales. The bottoms from the fractionator are recycled back to the absorber-stripper circuit as aqueous make-up.

The design and cost drivers for this section are:

- 1) Feed rate to stripper, lb/hr
- 2) Concentration of NH_3 in feed
- 3) Other constituents of feed, HCN, HCl, SCN, CO_2 , etc.

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 35
DESCRIPTION AMMONIA RECOVERY - CLC PROCESS

The CLC process, operated in conjunction with a Phenosolvan plant, has the proven capability of producing anhydrous ammonia from a dilute feed. Ammonia-containing aqueous streams are received from Phenol recovery, Unit Operation 34. Anhydrous ammonia of commercial purity (less than 10 ppm hydrogen sulfide) is produced and sent to By-Products Storage, Unit Operation 82, for sales. The CLC process is a proprietary process licensed by the American Lurgi Corporation.

The feed enters a stripper where the volatile components (ammonia and acid gas components such as carbon dioxide, hydrogen cyanide, and hydrogen sulfide) are separated from the condensate. The stripped condensate with less than 100 ppm free ammonia is sent to Biological Treatment, Unit Operation 37. A bleed stream of low boiling organic impurities goes to Incineration, Unit Operation 41.

The vapors or rich gas containing the volatile components are sent to the absorber where the ammonia is selectively absorbed in a circulating aqueous proprietary solution. The acid gases can be sent to Acid Gas Removal, Unit Operation 22. The ammonia-rich solution is heated, and the ammonia is reboiled out of the solution. The ammonia vapors are condensed and sent to By-Product Storage, Unit Operation 82, for sales. The remaining solvent-rich solution is recycled back to the absorber-stripper as aqueous make-up. The design drivers are:

- 1) Components of the feed
- 2) Concentration of ammonia in the feed.

The cost drivers are:

- 1) Concentration of ammonia in the feed
- 2) Feed rate to stripper, 1/hr.

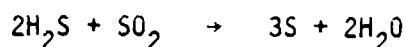
Raw materials needed included phosphoric acid and sodium hydroxide.

Critical technology areas needing further study and development work are:

- Contaminants - Data are needed relative to trace metals in the vapor phase streams of the system. The issue is the quality of the ammonia by-product.

SYSTEM NUMBER 15
UNIT OPERATION NUMBER 36
DESCRIPTION SULFUR RECOVERY - CLAUS SULFUR

Acid gas from Acid Gas Removal, Unit Operation 22, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. Typically in a Claus type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting the H_2S to elemental sulfur according to the following equation:



Any hydrocarbons in the acid gas are burned to CO_2 and H_2O .

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in Sulfur By-Product Storage, Unit Operation 83, and the tail gas from the final condenser is fed to a tail gas treating unit where substantially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere.

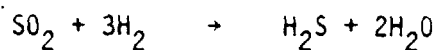
BEAVON TAIL GAS TREATING UNIT

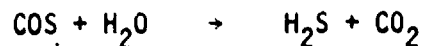
Several commercial processes are available for reducing the sulfur content of sulfur recovery unit tail gas to an environmentally acceptable level. The Beavon sulfur removal process is capable of reducing the sulfur content in the tail gas to less than 100 ppm. It is a proprietary process licensed by Ralph M. Parsons Co.

Hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled, and passed into a Stretford unit where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution, and transferred to Sulfur By-Product Storage, Unit Operation 83. The decanted redox solution is returned to the system.

The chemical reactions are:

Hydrogenation and Hydrolysis





Hydrogen Sulfide Conversion



The purified tail gas is odorless and contains typically less than one ppm of H_2S and less than 50 ppm of total sulfur compounds, mainly COS.

The sulfur product is yellow and better than 99.9% pure.

The Design Drivers for this system include:

- 1) Amount of sour gas available
- 2) Sulfur emission regulations
- 3) Hydrogen sulfide concentration of feed gas.

The Cost Drivers include:

- 1) Required system efficiency
- 2) Hydrogen sulfide concentration of feed gas
- 3) Fuel required in tail gas incinerators
- 4) Amount of steam generated.

Critical technology areas needing further study and development are:

- Acid gas minor components - The quantity of constituents such as HCN, COS, CS_2 , RSH, HxCx contained in the gas feed to the SRU needs to be established. The existence of these components effects the operability of the system, and the data would effect the system design.

SYSTEM NUMBER 18

UNIT OPERATION NUMBER 37

DESCRIPTION BIOLOGICAL TREATMENT - POAS PROCESS

The purpose of this unit is to treat effluents from sanitary waste treatment and the process condensate treating system, Unit Operation 33, to such an extent that they can be recycled into the facility water system.

This is a proprietary process licensed by Union Carbide Corporation.

This unit treats the various contaminated aqueous streams for recovery of recycled water. Process waste water from Ammonia Recovery is treated to remove any remaining ammonia, phenol and other organic compounds. The process waste water treatment plant consists of an equalization basin where wastewaters from various sources are mixed and the flow is equalized. The pH of the wastewater from the equalization lagoon is adjusted, and nutrients are added to satisfy the nutritional requirements of the microbiological mass responsible for bio-oxidation. The waste water then flows to an oxygen-activated sludge reactor where high-purity oxygen is supplied to the micro-organisms which metabolize organic materials. Due to the exothermic biological reaction, a portion of the mixed liquor from the biological treatment reactor is cooled in a heat exchanger and is then recycled to the treatment reactor. The rest of the mixed liquor is sent to clarifiers to separate the microbiological mass. The water then flows to Raw Water Treatment and Storage for recycle as process water. A portion of the activated sludge separated in the clarifiers is returned to the biological treatment reactor. Excess activated sludge will be wasted and pumped to Solid Waste Treatment, Unit Operation 31, for further dewatering and final off-site disposal.

The primary design and cost drivers for this unit are:

- Flow rate to treatment
- Quantity of contaminants
- Recycle water quality requirements.

SYSTEM NUMBER 1
UNIT OPERATION NUMBER 37
DESCRIPTION BIOLOGICAL TREATMENT - AIR ACTIVATED
SLUDGE PROCESS

Activated sludge waste water treatment is a biological oxidation process in which micro-organisms consume dissolved organic contaminants and convert these substances to carbon dioxide, water, and more micro-organisms. The activated sludge process consists of aerated biological reactors in which the influent waste waters are mixed with recycle sludge, and clarifiers which separate the biological sludge by quiescent settling, from the treated effluent. In the aeration basin, the sludge is referred to as mixed liquor suspended solids (MLSS). Clarifier effluent should contain about 25 ppm suspended solids, mostly biological solids. Excess sludge consisting of organisms in excess of those required to maintain steady-state biological concentrations in the reactors, are removed from the clarifier bottom recycle line.

Activated sludge biological oxidation is presently one of the most common methods for treatment of municipal and industrial wastes. It is a relatively simple process to operate particularly when the characteristics of the waste are consistent. In addition to organics, biological oxidation removes some amounts of trace metals, and trace organics. It also removes phenols, cyanides, and ammonia that are present in coal conversion wastes.

The system performance of the activated sludge is highly sensitive to process disturbances. It can be adversely affected by significant changes in process parameters such as pH, temperature, and organic and hydraulic loadings. Upset conditions could result in less acceptable effluent BOD and suspended solids levels.

Certain levels of nutrients should be maintained in the feed stream in order to satisfy nutritional requirements of biological agents responsible for bio-oxidation. Macro nutrients such as nitrogen and phosphorous are required in proportion to organic content of the feed. A typical weight ratio used by biological wastewater treatment is BOD: N: P = 100: 5: 1. While residual ammonia levels can be controlled in the stripped process, phosphorous levels should be maintained by adding chemicals such as H_3PO_4 . Micro nutrients such as managanese, copper, zinc, and other metals may also have to be maintained at proper levels.

In gasification processes, a part of the organic nitrogen contained in the feed coal will be converted to hydrogen cyanide (HCN). In gas cooling, Unit Operation 21, HCN is removed from the gas by quenching. HCN in the sour water is partly removed in Ammonia Recovery, Unit Operation 35. However, part of the HCN could react in the water feeding the biological treatment plant. It has been found in coke oven plants that the thiocyanate is more difficult to destroy than the feed organics and about twice the reactor residence time is required for its removal. More data is

needed to quantify the thiocyanate found in the process quench condensate and its effect on the performance of bio-oxidation unit.

The cost and design drivers for this unit are the same as for the POAS Process described earlier.

SYSTEM NUMBER 17
UNIT OPERATION NUMBER 39
DESCRIPTION COOLING WATER SYSTEM

The purpose of this unit is to provide cooling water to the various process users in the facility.

This is a non-proprietary system and is supplied by several US vendors.

Each cooling tower system would include the tower and fans, sidestream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

Typically, a portion of the circulating water is passed through side stream filters to reduce loading of suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basin is maintained by continuous make up of clean water from the raw water treatment system.

The blowdown stream contains inhibiting chemicals that are objectionable to release to the environment. The blowdown stream would be passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algecides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors.

The Design Drivers are:

- 1) Makeup Water Quality
- 2) Cycles of Concentration
- 3) Air Relative Humidity/Temperature.

The Cost Drivers are:

- 1) Capacity
- 2) Approach Temperature
- 3) Side Stream Treatment
- 4) Chemical Treatment Recovery.

Critical technology areas needing further study and development are:

- Make-up water - Treatment requirements required for use of coal-derived process waste water for cooling water makeup need to be determined.

SYSTEM NUMBER 9

UNIT OPERATION NUMBER 41

DESCRIPTION

INCINERATION

This unit combusts environmentally objectionable constituents in various vent gases and waste gases from facility systems and renders these gases into a form that is acceptable for release into the atmosphere.

This is a non-proprietary process.

This unit typically consists of a combustion chamber (burner) and a tall stack to disperse the combusted gases. The waste gas streams usually fed to this unit are lock hopper gases, other off-gases, or flue gas desulfurization exit gas and coal dust or fines.

This unit would not be required in a gasification facility using a TEXACO or KOPPERS-TOTZEK gasifier and with no coal fired boiler.

The Design Drivers include:

- 1) Gas Temperature, °F
- 2) Gas Fuel value.

The Cost Drivers include:

- 1) Capacity, Mlb/HR
- 2) Gas Properties
- 3) Incinerator Type
- 4) Heat Recovery.

No critical technology issues have been identified.

SYSTEM NUMBER 16

UNIT OPERATION NUMBER 80

DESCRIPTION

AIR SEPARATION AND OXIDANT FEEDING

The purpose of this unit is to supply oxygen to the gasifiers. The gaseous oxygen stream is usually between 95 and 98 percent pure by volume, and it is produced by distillation of liquified air.

This is a proprietary but non-licensed process offered by several designer/manufacturers in the U.S.

In a typical Air Separation unit, atmospheric air is filtered and compressed to the required pressure by a two-stage air compressor, driven by either steam turbines or electric motors, depending on the plant steam balance.

The compressed air is first cooled in water after-coolers and finally cooled in the switching passages of a reversing heat exchanger to just above its liquefaction temperature. As it cools, water and carbon dioxide freeze out on the walls of the exchanger and are thus removed from the air stream.

The water and carbon dioxide are removed from the exchanger walls by switching the exchanger passages to allow the waste nitrogen to flow through the former air passages. The waste nitrogen stream is at a low enough pressure such that sublimation of the water and carbon dioxide into the nitrogen is possible. This waste stream is vented to atmosphere.

The cold air then enters the high pressure column for initial purification. The vapor at the top of the column is nitrogen containing 10 ppm of oxygen. Most of this is condensed in the High Pressure Condenser/Low Pressure Reboiler, and returned as reflux to the High Pressure Column. Some of this nitrogen vapor becomes the feed stream to the expansion turbine and some becomes low pressure product nitrogen gas after being superheated in the Superheater and being warmed to about -150°F in a reversing heat exchanger.

Some of the nitrogen condensed in the condenser/reboiler becomes the liquid nitrogen product. It is first subcooled in the Waste Nitrogen Core Sub-cooler and then flashed into the Liquid Nitrogen Separator. Liquid nitrogen product from the separator is transferred to the liquid nitrogen storage.

An impure reflux stream containing a small percentage of oxygen is withdrawn at an intermediate point in the column. It is subcooled and flashed into the top of the Low Pressure Column.

The final purification of oxygen takes place in the Low Pressure Column. The feeds to this column are impure reflux, and crude liquid oxygen from the High Pressure Column which is subcooled and passed through the

Hydrocarbon Absorbers before being flashed into the Low Pressure Column. This procedure assures that all hydrocarbons are removed from the oxygen before purification. Processing the crude oxygen after it is liquified assures that the absorption is done at a relatively constant pressure and temperature and thus limits the possibility of accidental desorption.

As a further safeguard, a portion of the liquid oxygen in the low pressure sump is constantly withdrawn, passed through a guard hydrocarbon absorber and returned to the sump. This prevents trace quantities of hydrocarbons from accumulating in the low pressure sump.

Product oxygen gas and liquid is withdrawn from the bottom of the Low Pressure Column. The product oxygen gas is superheated in the superheater and then warmed to ambient temperature in the reversing exchangers. A four-stage centrifugal compressor is then used to compress the oxygen gas to the required pressure to feed the gasifiers.

The liquid oxygen product, a small fraction of the total oxygen product, is withdrawn from the Low Pressure Column sump, subcooled in the oxygen core subcooler and transferred to LOX storage.

The main cost and design drivers of this unit are:

- 1) Product Purity
- 2) Product Pressure
- 3) Plant capacity.

Product Purity requirements determine the number of trays and column diameter required to achieve the specified purity. Washer Product Purity requires larger, taller distillation columns.

Product Pressure requirements set by the gasifier pressure set the necessity for Product oxygen compressors. Low Pressure gasifiers do not require Product compression, where High Pressure gasifiers required the inclusion of multi-stage centrifugal compressors.

Plant capacity requirements set the overall equipment size and number of units. Usually 2 or 3 trains of air separation are required to provide 1500-2000 tons per day of oxygen.

Critical technology issues needing further study and development are:

- Compressor technology - Improved large compressors technology is needed for increased-size safe O₂ systems.

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 81
DESCRIPTION FINAL SOLIDS DISPOSAL

The purpose of this unit is to store solid waste generated by facility operation during the facility life.

This is a non-proprietary system.

This system consists of a lined impounding pit sized to contain 20 years of solid waste. It typically includes the conveyor system to move the solids to the pit and a leachate recovery area to recover and pump leachate to the process condensate system.

The primary cost and design drivers are the quantity of solids to be disposed and the volume of the pit to be excavated and lined.

Critical technology issues needing further study and development are:

- Ground water quality - The issues are leachability of solids, monitoring of ground water, and control of runoff.

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 82
DESCRIPTION BY-PRODUCT STORAGE AND LOADING

The purpose of this system is to receive and store various by-products from recovery units, transfer the by-products to a loading area and into the proper vehicles for transportation.

This system will receive and store: (1) ash residue from Coal Preparation and Feeding, Unit Operation 11; (2) oils and tars from Tar-Oil Separation, Unit Operation 32; (3) liquid anhydrous ammonia from Ammonia Recovery, Unit Operation 35; and (4) phenol from Phenol Recovery, Unit Operation 34. All tankage will be conventional large diameter low pressure tanks with fixed roofs except for the pressure spheres which will store the ammonia by-product. Transfer pumps will send the by-products to the loading area for filling the vehicles used to deliver the by-products to sales. This is a non-proprietary system.

The Design Drivers for this system include:

- 1) Number of by-products
- 2) Quantity stored
- 3) Quantity and type of shipping.

The Cost Drivers include:

- 1) Number of by-products
- 2) Loading type
- 3) Length of storage needed.

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 83
DESCRIPTION SULFUR BY-PRODUCT STORAGE

Sulfur of high purity from Sulfur Recovery and Tail Gas, Unit Operation 36, flows by gravity to sulfur pits. From the pits it is pumped as a 280°F liquid to storage tanks awaiting loading on a barge or truck. This is a non-proprietary system available from a variety of U.S. vendors.

The tanks are large diameter, low pressure of mild steel construction with capacity in excess of 10,000 barrels.

Typically, the tanks are insulated and steam traced on the shell, roof, and bottom and equipped with a sulfur pump.

The pumps, piping, and loading arms will be insulated and steam jacketed. Loading meters are provided to facilitate sulfur loading and to ensure inventory control.

The design driver for this system includes the ambient temperature range (heat loss from tankage).

The cost drivers include:

- 1) Moles per hour of sulfur recovered in Unit Operation 36
- 2) Hours of storage required.

No critical technology issues have been identified.

SYSTEM NUMBER 15

UNIT OPERATION NUMBER 84

DESCRIPTION

STEAM GENERATION

The purpose of the steam generation unit is to provide high pressure superheated steam to supplement the steam generated by waste heat recovery in the Gasifier Unit Operational Cooling Systems, Unit Operation 21.

The HP steam generators are water tube boilers fired either by raw coal, coal fines or MBG.

This is a non-proprietary process, and the equipment is supplied by several U.S. vendors.

An MBG Boiler (HP steam generator) is typically a balance draft, industrial type. It would include the water tube boiler, superheater, economizers, forced draft and induced draft air preheaters, and burners suitable for MBG.

Deaerated boiler feed water (BFW) is pumped through economizing coils, the water tube boiler, superheating coils and into the HP steam distribution system. Air for the combustion of MBG is supplied by a forced draft fan and heated in exchange with boiler flue gas which is supplied by the induced draft fan.

A coal fired HP Boiler is similar to the MBG Boiler with the exception of the ancillary equipment necessary for transport, storage, and combustion of coal or coal fines.

Typically, coal or coal fines are transferred to the boiler area by a conveyor which distributes the coal or coal fines to HP boiler coal storage bins. This is fluidized by primary air fans and transferred to the burners. Secondary air supplied by the forced draft fan recovers heat from the flue gas leaving the boiler to increase the efficiency of the boiler operation. Flue gas is transferred to the flue gas desulfurization system, Unit Operation 86, by induced draft fans before release to the atmosphere.

Chelating agents (Ethylene Diamine Tetra Acetic Acid) is added to boiler to reduce dissolved solid carryover into the HP steam. Di and Tri Sodium Phosphate (or polyphosphates) and filming amines (octyldecylamine) are added to the HP steam to control pH and reduce scaling problems.

The Design Drivers include:

- 1) Steam Requirements
- 2) Coal Properties.

The Cost Drivers include:

- 1) Capacity, Mlb/HR
- 2) Steam Pressure, PSIG
- 3) Degree of Superheat
- 4) Flue Gas Treatment Req'd
- 5) Boiler Type.

Critical technology issues needing further study and development are:

- Combustion of by-products, tar and crude phenol is a means of utilizing these products. Demonstration of combustion technology for these products is needed.

SYSTEM NUMBER 16
UNIT OPERATION NUMBER 85
DESCRIPTION RAW WATER TREATMENT

The Raw Water Treatment Unit is typically designed to provide treated and untreated water for the following facility water systems:

- Fire Water
- Service Water
- Potable Water
- Cooling Water
- Boiler Feedwater

The processes by which the raw water is treated for the above services are non-proprietary and are supplied by several US vendors.

Raw water is usually pumped from the river to a Fire Water-Raw Water Storage tank (or pond).

The Raw Water-Firewater Storage Tank provides surge capacity for Water Treatment as well as storage capacity for firewater. During an emergency, firewater is pumped from the tank to the firewater header system. The firewater pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the Raw Water-Firewater Storage Tank to the Softener-Clarifier. Lime, alum, and polyelectrolyte from the Clarifier Bulk Chemical Storage and Feed System are added to the Softener-Clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one percent by weight sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the Softener-Clarifier flows to the Self-Backwashing Sandfilters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the Sandfilter Backwash Sump and is recycled to the Softener-Clarifier. The filtered water flows to the Filtered Water Storage Tank and is utilized as cooling tower make-up for the Process Cooling Tower, service water for general plant use, feed to the Demineralizer Package, and feed to the potable water system.

Water intended for potable services is usually chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored

in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 PPM free chlorine in the tank.

Filtered water intended as feed to the Demineralizer Package is usually injected with Sodium Sulfide to remove trace amounts of chlorine which adversely affect the Demineralizer resins and often filtered through activated carbon to remove any remaining organic contaminants and dissolved iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong exchanger is also provided. The cation exchangers remove cations such as calcium cation and magnesium, while the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

A mixed bed polisher is usually provided to remove silica to 0.02 PPM and to polish returned turbine condensate for reuse.

The demineralized water and condensate is preheated to 220-225°F before being pumped into the boiler feedwater deaerators.

Acid, stored in the Acid Tank, is used to regenerate the cation exchangers. Caustic, stored in the Caustic Tank, is used to regenerate the anion exchangers.

Both the Acid and Caustic tanks have pumps and metering devices to control the flow of regenerants to the demineralizer package.

The Boiler Feedwater Deaerating heaters operate at 15 psig and 250°F. The Deaerators reduce the oxygen content of BFW to 0.005 cc/lite.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feedwater pumps to protect the condensate systems.

The Design Drivers include:

- 1) Steam Pressure, PSIG
- 2) Raw Water Quality.

The Cost Drivers include:

- 1) Capacity, GPM
- 2) Water Treatment Level
- 3) Metallurgy
- 4) Ion Exchange Resins
- 5) Reverse Osmosis Elements.

No critical technology issues were identified for this system.

SYSTEM NUMBER 15
UNIT OPERATION NUMBER 86
DESCRIPTION FLUE GAS TREATMENT - WELLMAN-LORD
SO₂ RECOVERY PROCESS

Effluent gas containing sulfur dioxide and particulates from Steam Generation, Unit Operation 84, is treated in this system using the Wellman-Lord sulfur dioxide recovery process. This sulfur dioxide recovery system includes absorption equipment, chemical plant for regeneration, and purge treatment plant. This is a proprietary process licensed by Davy Powergas, Inc., Houston, TX.

The system is designed to process gas containing sulfur dioxide and fly ash. The flue gas is picked up from the Steam Generation System, Unit Operation 84, by hot-side fans. The discharge pressure from each fan is sufficient to force the flue gas through a precipitator-type prescrubber and tray-type sulfur dioxide absorption tower. The cleaned gases from the absorbers are reheated before discharging to the atmosphere.

The sulfur dioxide-rich absorbing solution is processed in the regeneration area.

Chemical regeneration uses forced circulation exchangers. The rich absorbing solution is thermally regenerated by driving off sulfur dioxide and water vapor. This vapor is concentrated, compressed, and forwarded to Sulfur Recovery, Unit Operation 36. Condensate containing removed particles is sent to Solids Treatment System, Unit Operation 31. The regenerated absorbing solution leaving the evaporators is combined with stripped condensate from the partial condensers and recycled back to the absorbing solution storage area.

A slipstream of absorbing solution is taken from the absorbers of the power plant for processing in a purge treatment area. Through a series of unit operations, the inactive sodium salts are concentrated, removed from the remaining absorbing solution, and sent either to By-Product Storage, Unit Operation 82, or to Solids Disposal, Unit Operation 81.

The Design Drivers for this system include:

- 1) Concentration of SO₂, and SO₃ in flue gas
- 2) Concentration of HCl, ash, and other constituents
- 3) Concentration of oxygen in flue gas
- 4) Superficial velocity of gas in absorber, ft/sec.

The Cost Drivers include:

- 1) Percent sulfur dioxide removal required
- 2) Gas flow to absorber, ACFM
- 3) Moles per hour of SO_2 and SO_3 removed
- 4) Treated gas reheat requirements, i.e., the difference in temperature between gas from absorber and gas to stack.

By-products include:

- 1) Sodium bicarbonate
- 2) Sodium bisulfite
- 3) Sodium sulfate.

Critical technology issues needing further study and development are:

- Effect of gas contaminants such as HCl on materials of construction
- Mist elimination technology
- Demonstration of long-term operation at high SO_2 removal efficiency.

SYSTEM NUMBER 15
UNIT OPERATION NUMBER 86
DESCRIPTION FLUE GAS TREATMENT - DOUBLE ALKALI

The purpose of this unit is to absorb SO_2 from flue gas in a scrubber using a sodium carbonate solution.

This proprietary process technology is offered by several U.S. vendors, including Combustion Engineering Associates and FMC Corp. In the sodium double-alkali process, flue gas from Steam Generation, Unit Operation 84, at a temperature of about 350-400°F, is compressed through a pressure rise of about 10" water by booster fans, and then enters an Absorber. In the Absorber, the gas passes upward through a series of sprays, a set of trays, and then through a demister. The clean flue gas leaving the Absorber is reheated indirectly by 50 psig steam to about 190-200°F and leaves the facility through a tall stack.

The gas in the Absorber is countercurrently contacted as it passes through the trays, by an aqueous solution of NaOH , Na_2SO_3 , Na_2SO_4 and Na_2CO_3 . The SO_2 rich solution is collected at the bottom of the Absorber and recycled to the tray and spray sections. The absorbent reacts with the SO_2 in the flue gas to produce a mixture of sodium salts.

A continuous bleed stream of liquor is sent to the Absorbent Regeneration Subsystem. The absorber bleed first goes to a Surge Tank from which it is pumped to Reaction Tanks. A slipstream is routed to pick up quicklime brought into the unit from outside and is then pumped back to the Reaction Tanks.

The Reaction Tanks are designed to convert the soluble sodium salts into less-soluble calcium sulfite, which precipitates and results in a slurry containing about 5 percent solids by weight. The slurry is then pumped to a Thickener.

Overflow from the Thickener is pumped back through a Holding Tank to the Absorber. Soda ash is added to a slipstream to replace sodium lost from the system with the filter cake.

Underflow from the Thickener contains about 25 percent solids by weight, and is pumped to vacuum filters where the solids are dewatered to a cake containing about 60 percent solids by weight. Washing of the cake causes most of the sodium salts to be dissolved in the filtrate, which is returned to the Absorber.

The solids cake from the Vacuum Filters is conveyed to Final Solids Disposal, Unit Operation 81. The cake has excellent handling properties due to the precipitation of mixed crystals of $\text{CaSO}_4/\text{CaSO}_3$. The material is non-thixotropic, drains well, and does not reslurry when exposed to rain. Since, however, the cake will contain volatile trace elements produced by

combustion of char and scrubbed from the flue gas in the flue gas desulfurization unit, and will also contain NaCl produced from HCl in the flue gas, the solids must be disposed of in a secure onsite landfill.

The primary cost drivers are:

- 1) ACFM to the absorber
- 2) % SO_2 removal

The primary design drivers are:

- 1) PPM SO_2 - SO_3 in flue gas
- 2) % HCl; ²ash, etc., in flue gas.

Critical technology issues needing further study and development are the same as those for the Wellman-Lord Process, plus better data on the distribution of toxic trace elements between the flue gas and the waste solids.

SYSTEM NUMBER 14
UNIT OPERATION NUMBER 87
DESCRIPTION PLANT ELECTRICAL SYSTEM

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

- Develop the necessary voltage stepdown arrangement for plant requirements
- Distribute the necessary power to the plant equipment.

This is non-proprietary equipment and is supplied by several U.S. vendors.

TVA's incoming substation transformers receive power from its prevalent distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirements for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution System is typically a secondary selective system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Typical load center transformers are air cooled, dry type, 150 F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consists of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 Volt DC System supplies control power for medium voltage and 480 V volt plant switchgear control, protective relaying and annunciation. The system also supplies power for emergency lighting.

The Design Drivers include:

- 1) Reliability Factor
- 2) Motor Horsepower.

The Cost Drivers include:

- 1) Capacity, KW
- 2) Voltage Level, KV
- 3) Peak Current Demand, AMPS
- 4) No. of Services.

SYSTEM NUMBER 19
UNIT OPERATION NUMBER 88
DESCRIPTION BUILDINGS AND SUPPORT FACILITIES

The purpose of this unit is to provide equipment or services to support the Gasification Facility at the facility level.

The equipment and services provided in this unit are non-proprietary.

This unit is a general facility category and would typically provide the following equipment or service:

- Administration Building
- Laboratories
- Change Rooms
- Warehouses
- Maintenance Buildings
- Operation Centers
- Security Offices
- First Aid Facility
- Fire House
- Visitor Reception
- Plant Fencing
- Plant Lighting
- Roads, Bridges, and Sewers
- Docking Facilities
- Interconnection Pipe Ways
- Fire Protection Network
- Flare Stacks and Headers
- Plant Instrument Air Compressors
- Environmental Monitoring
- Site Preparation.

The primary cost and design drivers are the number of and the extent of each of the above items provided for the facility.

SYSTEM NUMBER 10
UNIT OPERATION NUMBER 89
DESCRIPTION CONTROL AND INSTRUMENTATION

The purpose of this unit is to provide operational control of the facility and supervisory master control of the facility module operation.

This item has not been identified as a discrete cost or operational center in the studies or evaluation presented in the literature.

A separate subcontract is often executed with a supplier of computer data acquisition systems, to provide a data acquisition system for a facility and to interface it with the instrumentation and control loops within the facility.

2.4 System Stream Characterization and Raw Materials

Each unit operation within a coal gasification facility may be described in terms of inputs, outputs, raw materials necessary to support the process, and potential by-products. Appendix C is a summary of these items for the unit operations described in Section 2.3.

Minimal numerical data was reported with the unit operations because feed and effluent stream composition vary from plant to plant and reflect conditions not always reported in literature documents. The primary reason for this is non-disclosure agreements.

2.5 System Characterization

The characterization of each unit operation has been based on quantifying the cost and design drivers of each of the units.

Reference sources that are applicable to the individual unit have been reviewed and certain data extracted. Not all units are well represented in the data base and not all references were used in reporting data.

The characterization found in Appendix B of this report identifies the cost and date of monies for each system and quantifies appropriate cost and design drivers. Some cost and design drivers are well represented with quantitative data. Some, which are process design or client specific, have been identified but without corresponding numerical data.

2.6 System Component Characterization

Each unit operation contains a discrete and, generally, specific set of components. The components can be segregated into 13 broad categories, which have usually been used by A/E's to prepare the semi-definitive cost estimate of a unit or facility. Depending on the unit design and purpose, some of these components can affect project costs and scheduling and must be handled accordingly. Some components, because of size, metallurgy or complexity require long times to deliver to the job site or higher than usual costs. Table 2-3 identifies the usual component breakdown of each system and identifies those equipment items which typically require long lead time, have high costs, or are items of critical technology.

2.7 Operations and Maintenance Costs

Preparation of facility cost estimates and life cycle costing requires an estimation of facility operating and maintenance costs. The costs can be broken down on a unit operation basis, but they are usually estimated based on total capital investment of the facility. A common estimation practice is to represent the various costs elements in this category as a percentage of the equipment installed capital cost. Maintenance expense is usually estimated as 1 to 6 percent of capital investment with a 60/40 material to labor split.

TABLE 2-3. SYSTEM COMPONENT CHARACTERISTICS

	FIELD FABRICATED VESSELS	SHOP FABRICATED VESSELS	COMPRESSORS BLOWERS	EXCHANGERS	PUMPS	FIELD EQUIPMENT	SOLID TRANSFER	SOLID STORAGE	FILTRATION	CLASSIFICATION	SPECIALTY EQUIPMENT	SIZE REDUCTION CLASSIFICATION	LABOR
10 COAL RECEIVING, STORAGE & TRANSFER & FEEDING	0	4	4	0	4	0	7	4	0	0	1, 2	0	0
11 COAL PREPARATION & FEEDING	0	1, 3, 4, 11	4	0	0	4	7	4	0	0	4 (OR 7)	0	0
20 GASIFICATION (NOTE 1)	0	3, 3, 3, 3	0, 2, 4, 0	0, 2, 0, 0	4	0, 0, 0, 0	14, 4, 4, 4	0, 0, 0, 4	0, 0, 0, 0	0, 0, 0, 0	0, 0, 0, 0	0	0
21 GAS COOLING (NOTE 1)	0	4	0	4, 3, 4, 3, 3	4	0	0	0	0	0	0	0	0
22 ACID GAS REMOVAL (NOTE 2)	1, 4, 4, 4	1, 2, 3, 1	1, 1, 0, 4	1, 0, 1, 1	1, 4, 4, 4	0	0	0	4, 4, 4, 4	0	0	0	0
23 COMPRESSION	0	0	7	0	0	0	0	0	0	0	0	0	4, 4, 4, 4
31 SOLIDS TREATMENT SYS	0	0	0	0	0	0	0	0	0	0	0	0	0
32 TAR OIL SEPARATION	0	4	0	4	1, 2	0	3	0	0	0	0	0	0
33 PROCESS CONDENSATE TREATMENT	0	1, 2, 3	0	0	4	0	0	0	3, 2	0	2	0	4
34 PHENOL RECOVERY	0	0	0	7, 3	2	0	0	0	0	0	4	0	0
35 AMMONIA RECOVERY	0	1, 2, 3	0	4	4	0	0	0	0	0	0	0	0
36 SULFUR RECOVERY	0	1, 2, 3	0	4	4	0	0	0	0	0	1, 2, 3	0	0
37 BIOLOGICAL TREATMENT	0	0	0	0	2	0	0	4	0	0	1, 2	0	0
38 COOLING WATER SYSTEM	0	0	0	0	4	0	0	4	0	0	0	0	0
41 INCINERATION	0	0	0	4	4	0	0	0	0	0	0	0	0
49 AIR SEPARATION AND OXYGEN FEEDING	0	0	4	4	0	0	0	0	0	0	0	0	0
41 FORMAL SOLIDS DISPOSAL AND FEEDING	0	0	0	4	0	0	0	0	0	0	0	0	0
42 BY PRODUCT STORAGE AND FEEDING	0	0	0	0	3	0	1	0	0	0	0	0	1, 2
43 SOLIDS STORAGE AND FEEDING	0	0	0	0	4	0	0	0	0	0	0	0	0
44 STEAM GENERATION	0	0	0	0	1	0	0	0	0	0	0	0	0
45 RAW WATER TREATMENT	0	0	0	4	4	0	0	4	0	0	0	0	0
46 FINE GAS TREATMENT	1, 2, 1	1	1, 2	1	2, 3	0	4	4	0	0	1, 2, 3	0	0
47 PLANT ELECTRICAL SYS	0	0	0	0	0	0	0	4	0	0	1, 2	0	0
48 TAILING & GENERAL STORAGE FACILITIES	0	0	0	0	4	0	0	0	0	0	1, 2	0	0
49 CONTROL & INSTRUMENTATION (NOTE 3)	0	0	0	0	4	0	1	0	0	0	1, 2	0	0

CHARACTERIZATION LEGEND: 0 NOT INCLUDED IN SYSTEM
1 HIGH CYCLE ITEM
2 LONG LEAD TIME ITEM
3 CRITICAL TECHNOLOGY AREA
4 NON CRITICAL ITEM

NOTES: (1) NUMERICAL SUBSCRIPTS REPRESENT THE FOLLOWING: ORDER OF VASHERS, MAX. THERMAL & T. LONG. INCINERATOR
(2) NUMERICAL SUBSCRIPTS REPRESENTS THE FOLLOWING: ORDER OF AIR UNITS, STEAM, STRIPPING, REIFIED REFINISH
(3) NOT INCLUDED IN SOURCE REFERENCE LITERATURE, NO DATA REPORTED.

For this reason, the references that were most applicable to a coal gasification project were reviewed, and the quantitative operating and maintenance costs were reported as a function of total installed capital.

THE SDM CORPORATION

3.0 RAW MATERIAL ANALYSIS

3.0 RAW MATERIALS ANALYSIS - TASK 5.1.2

Most of the unit operations identified in this study require raw materials to support the process function. These raw materials are necessary for unit performance and represent a cost of operating both annually and as an initial cost.

This section presents the results, summarized in Table 3-1, of Task 5.1.2. The approach taken is to present all raw material elements that are required by all the gasification facility system candidates that were identified in Task 5.1.1. Therefore, dependent upon the final design configuration, only some of the raw material listed in Table 3-1 will be required to support the selected unit operations for the TVA Coal Gasification Facility.

The data in Table 3-1 is presented in a matrix format. It includes a list of raw materials, the system(s) with which they are associated, both the initial and replacement quantities required, the unit cost, a commercial source, and transportation information.

The list of raw materials and the system(s) with which they are associated were obtained during Task 5.1.1. It should be noted that some items in the raw materials list (e.g., Strong Acid Cation Resin) represent a general class of materials and, of necessity, the cost listed represents a specific material in that class. The material chosen for such cases was the one that is often used in applications such as this and has a representative cost. However, the specific material selection must await final system design. The initial and replacement quantities are approximate amounts based upon information from reference literature and vendors and upon prior design experience and engineering judgment. The exact quantities required will be obtained during final system design. The unit cost and transportation information was obtained from the listed commercial sources and represents current market conditions.

The data presented in Table 3-1, while not exact at this point in the program, is sufficiently accurate to be used as a basis for determining the operating costs of reference facilities to be used in the comparative evaluations of the A/E designs. Based on the final design chosen, the data in Table 3-1 will be further refined and updated during Task 5.4.2.2 and will be used in Task 5.4.2.4 to develop life cycle cost estimates.

TABLE 3-1. RAW MATERIALS SUMMARY

[illegible]

THE BDM CORPORATION

4.0 COAL GASIFICATION BY-PRODUCT MARKET ANALYSIS

4.0 COAL GASIFICATION BY-PRODUCTS MARKET ANALYSIS

4.1 Introduction and Summary

This section describes a market analysis of the by-products from the TVA Coal Gasification Facility located in Northern Alabama. Since the plant is not conceptually designed, and the by-products are dependent on process type and design, quantities of the by-products are not included in this report. The generic price, use, and future market expectations are shown in Table 4-1 for the by-products considered in this analysis. Oxygen is a possible export because all modules will not become operational at the same time and all the oxygen produced will not be consumed until the entire facility becomes operational. However, because of the short-term nature of this situation, oxygen from this plant must be considered "merchant" oxygen rather than tied to a long-term contract. Oxygen, nitrogen, argon and carbon dioxide are treated as gases, and transportation by pipeline or truck is not considered. Tar, slag, and ash contents are not defined, therefore, they are treated generically. Sulfur is an excellent by-product candidate for marketing.

TABLE 4-1. BY-PRODUCT SUMMARY

BY-PRODUCT	U.S. POTENTIAL USE	CURRENT PRICE (EARLY 1980)	DISTANCE FROM PLANT WHERE TRANSPORTATION COSTS=CURRENT PRICE (MILES)	TVA MARKET PENETRATION RATING
SLAG/ASH	ROAD AND CEMENT FILL, INSULATION, FLUE GAS DE- SULFURIZATION, LAND FILL	NONE (80% DUMPED)	-----	POOR
SULFUR	PHOSPHATE FERTILIZERS, PAPER, SOIL NUTRIENT, ROAD BASE	\$116/LONG TON + \$30 (SPOT)	> 1500	GOOD
OXYGEN	METAL MANUFACTURING, HEALTH SERVICES, METAL FABRICATING	34¢/100 FT ³ & SHIPPING & EQUIPMENT	-----	FAIR
NITROGEN	BLANKETING ATMOSPHERES- CHEMICALS, ELECTRONICS, METALS, FREEZING AGENT, AEROSPACE	30.5¢/100 FT ³ & SHIPPING & EQUIPMENT	-----	FAIR
CARBON DIOXIDE	FOOD REFRIGERATION INDUSTRIAL REFRIGERATION, CARBONATION	\$60/TON	1100	GOOD
TAR	CHEMICALS, FUEL	\$95/TON	> 1500	GOOD
STEAM	PROCESS HEAT	\$5-\$6/10 ⁶ BTU	< 10	FAIR
ARGON	LAMP FILLER, WELDING CUTTING, CRYSTAL GROWTH	73.2¢/100 FT ³	-----	FAIR

Conclusions of by-product market analysis are:

- Quantifying TVA region market penetration can only be done if quantity and specifications of by-products are known.
- Gases will have to be liquefied or transported by pipeline over relatively short distances (liquefaction costs have not been determined).
- Most of the by-products will probably be used by new industries locating near the Murphy Hill site.
- TVA design criteria specifies only truck and barge transportation, so rail costs are not included.
- Slag and ash would probably be impounded unless environmental constraints on utilization are relaxed.
- Tar could be converted by processing to fuel for industries during natural gas curtailment. This may be expensive and may lower the price which purchasers are willing to pay for the tar.
- Sulfur is a prime by-product candidate for marketing.
- Gases will probably have to be purified for market sales.

4.2 Transportation

TVA design criteria specifies only truck, barge, and pipeline transportation modes are to be considered in the facility conceptual design. From Reference 1, the following transportation costs (1980) were used in the analysis; tariffs for interstate transportation is not included.

TRUCK: \$5.61/Ton/100 Mile

BARGE: \$.0088/Ton Mile (> 40 Miles)

Plus 18¢/Barrel for One Terminal Charge.

Truck transportation is the most flexible mode of shipping, with almost half of all petroleum products currently being shipped by truck. Most trucking is done by contract carriers and the prices included in this analysis are for distances greater than 100 miles. Upper limits of petroleum-product truck hauling are around 300 miles.

Barges currently ship a small percentage of the total petroleum products in the United States. Barge costs are relatively inexpensive, including terminal charges of approximately 18¢/barrel. It is most economically feasible to transport by-products by barge greater than 200 miles. The 8.8 mils barge cost per long ton mile are escalated from the data reported in Reference 1. Figure 4-1 shows by-product transportation costs for barge.

Until the 1960's, elemental sulfur was handled and shipped in solid form. However, at present about 90 to 95 percent of the elemental sulfur consumed in the United States remains as a molten liquid from the point of production (the mine in the case of Frasch sulfur or the plant in the case of recovered sulfur) to points of consumption at customers' plants. The reasons are convenience and economics of handling and shipping. Because most consumers use sulfur in liquid form, the need for remelting is eliminated. In addition, liquid sulfur is less likely to be contaminated during transportation and in storage.

The molten sulfur is transported in heated and insulated barges, ships or rail tank cars, maintained at a temperature above 238°F, and delivered direct to terminals or customers' heated storage tanks. From there, it is pumped directly to the sulfur burners of sulfuric acid plants or other sulfur using facilities. While the recovered sulfur producers depend most on direct shipment to consumers, the Frasch producers distribute their product largely through an extensive system of liquid sulfur terminals located in sulfur-consuming market areas. Some terminals are producer-owned, while others are consumer-owned. The Frasch industry also maintains a system of liquid and bulk sulfur terminals in Europe.

For storage purposes, molten sulfur is sprayed into vats to solidify in thin layers progressively one above the other, producing blocks of solid sulfur, which can later be broken up for shipment or remelting.

Solid or dry sulfur is sold in various forms including lump, granular, ground in various sizes or other special solid forms. It is shipped in bags or barrels or in bulk in railroad gondola cars, boxcars, highway trucks, barges or ships.

Hopper trucks are used to transport by highway a wide variety of materials. Vehicle types range from the open-dumping kind to the closed type. Most common is one that unloads by pressure differential into its own pneumatic-conveying system, which is temporarily connected to a storage silo. On this type of truck, the unloading of 40,000 lb. of products takes about 1 hour, sometimes less.

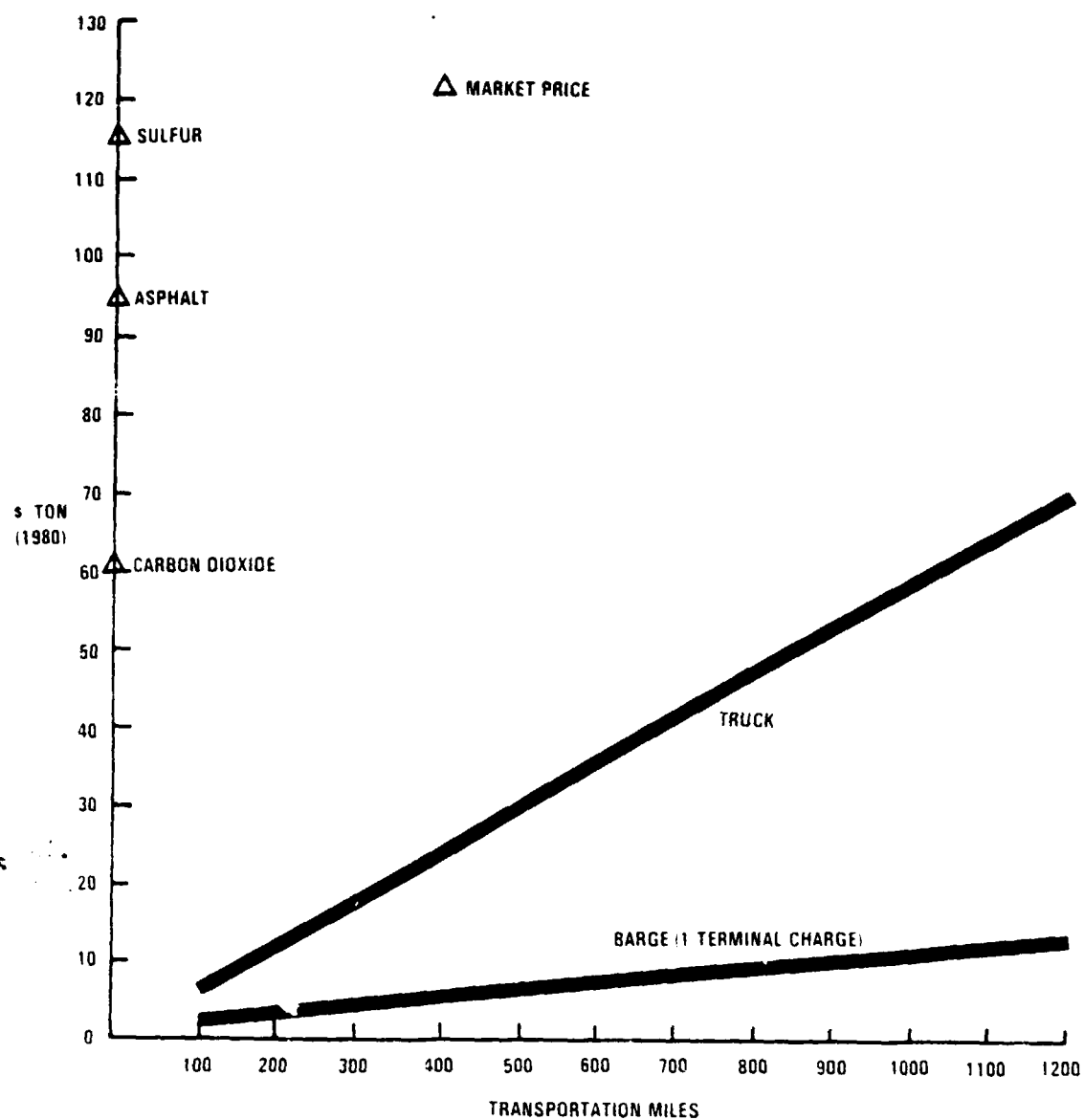


Figure 4-1. By-Product Transportation Costs

The actual weight that the truck can carry depends on the state-highway load limits, which in turn depend on the net vehicle weight and the number of axles on the truck (and tractor, when a trailer arrangement is used). The accepted maximum combined total weight of vehicle and cargo is 73,000 pounds. In some states, this is reduced slightly, while in others it is exceeded.

4.3 By-Products Cost and Market Assessment

The following subsections contain a brief analysis of each of the by-products in Table 4-1.

Slag and Ash

Fly ash, bottom ash, and boiler slag are considered in this analysis. Industrial coal ash, in general, has become a difficult problem for usage or disposal because the EPA may categorize coal ash as a toxic or hazardous waste. Under provisions of the Resource Conservation and Recovery Act of 1976, coal ash may be considered hazardous because of possible heavy-metals content. If EPA considers ash to be hazardous, facility designers will have to design collection and ponding systems similar to the mining industry. Disposal ponds would have to be impermeable to prevent leaking into aquifers.

Coal ash is currently the fifth most plentiful mineral substance produced in the U.S. (over 68 million tons per year). By 1985, the U.S. will be disposing over 100 million tons per year. Currently about one-fourth is being used as fillers for cement and roads.

Table 4-2 (Reference 2) summarizes coal ash collected and utilized from 1975 through 1978.

From its fossil fuel plants, TVA is currently disposing of over one million tons per year of ash and slag. TVA was attempting to use slag for rock-wool insulation by using a spin fiber technique. However, it was discovered that concrete blocks made with phosphoric-rock waste from the Muscle Shoals National Fertilization Development Center emitted low level radiation. TVA then put a hold on using any of its waste material, including fly ash and slag.

Current research underway in the U.S. for fly ash and slag utilization are (from Reference 2) the following:

- Fabricate fly ash based bricks.
- Recovery aluminum, iron, and titanium from ash and slag.
- Use fly ash for flue gas desulfurization.
- Form fly ash with moisture and calcium hydroxide to produce cement-like material.

TABLE 4-2. COAL ASH SUMMARY

	QUANTITY, 10 ⁶ TONS			
	1975	1976	1977	1978
<u>Ash Collected</u>				
Fly Ash	42.3	42.8	48.5	48.3
Bottom Ash	13.1	14.3	14.1	14.7
Boiler Slag	4.6	4.8	5.2	5.1
TOTAL	60.0	61.9	67.8	68.1
<u>Ash Utilized</u>				
Fly Ash	4.5	5.7	6.3	8.4
Bottom Ash	3.5	4.5	4.6	5.0
Boiler Slag	1.8	2.2	3.1	3.0
TOTAL	9.8	12.4	14.0	16.4
<u>Percent of Ash Utilized</u>				
Fly Ash	10.6	13.3	13.0	17.4
Bottom Ash	26.7	31.5	32.6	34.0
Boiler Slag	40.0	45.8	60.0	58.8
AVERAGE	16.4	20.0	20.7	24.1

There are different strategies for disposing of bottom ash, fly ash, and flue gas desulfurization (FGD) sludge at a coal gasification facility. Of course, waste disposal (or utilization) must be in an environmentally acceptable manner, taking into account capital and operating costs, and reliability. Currently, there are several types (Reference 3) of systems for handling coal wastes:

- Direct ponding
- Landfill (without treatment)
- Fixation/stabilization
- Forced oxidation/gypsum production
- Blending with fly ash.

The optimum choice of coal waste utilization will depend on the economics, environmental regulations, site terrain, and the process design.

In summary, coal ash, slag, bottom ash, and FGD sludge conclusions are:

- A 20,000 TPD facility will produce around 500,000 tons per year of coal waste which will probably be dumped on-site in a disposal pond.
- TVA is faced with increasing problems in disposing of coal waste because of environmental restrictions.
- Slag and bottom ash may have to be separated in order to find a commercial outlet.
- Detailed specifications regarding chemical and mineral content and the physical characteristics of ash products are basic determinants of marketability.
- Some ash is collected from coal burning facilities and used by commercial firms at no cost to the utility.

Sulfur

Sulfur is a very widely used industrial raw material. Production growth and demand has been increasing in the U.S. since the 1930's. Most sulfur consumption is in the form of sulfuric acid to which 85 percent or more of elemental sulfur is converted. Phosphate fertilization manufacture accounts for about 60 percent of sulfur consumption in all forms in the U.S., while the balance is in a wide range of applications in every sector of industry.

Sulfur as a volume commodity chemical is unique in the extent to which it is produced as a by-product or waste product derived from another primary product sought out and produced for its own worth. The by-product or waste product output may be considered "non-discretionary" in the sense that its production volume is not governed by its own demand. The second type, or primary product, is commonly termed "discretionary" because its volume of production is regulated directly by demand and price factors. Recovered sulfuric acid from oil and gas production, or burning fossil fuels or output of by-product sulfuric acid from metal smelting is "non-discretionary," while Frasch and pyritic sulfur are "discretionary." Because sulfur may be stored in solid form if it cannot be sold, its sales are not tied to its recovery, making it a "semi-discretionary" product.

A steady increase in demand, and a reduction in exports by foreign countries has currently created a tight sulfur supply situation both in the U.S. and worldwide. Spot shortages (Reference 4) will likely develop this year, and the tightness is not expected to ease before 1985. Sulfur production will be boosted by 2,200,000 long tons per year by 1982. After 1985, a temporary sulfur glut may occur due to expanded use of coal. In recent years, sulfur demand has grown due to growth in the key market of fertilizers. Sulfur demand growth is projected to be 3-5% per year during the early 1980's (Reference 4). Other traditional markets for sulfur, titanium dioxide, pulp and paper, and carbon disulfide, are expected to ease off by the late 1980's. New demands for sulfur may increase, such as using sulfur for replacement of asphalts or as a soil nutrient.

Sulfur-use growth will primarily depend on fertilizer demand. Fertilizer demand was booming in 1979, and is expected to continue strong in the early 1980's. However, fertilizer's history of big ups and downs lagging the general business cycles indicates that sulfur will have more cycles in the future. The Iranian crisis and fertilizer cutoff to the Soviet Union are also going to have major impacts on sulfur demand.

Total U.S. outlook for sulfur (Reference 4) is shown on Table 4-3 for 1979, 1980, and 1985.

TABLE 4-3. U.S. SULFUR OUTLOOK

PRODUCTION	MILLION METRIC TONS		
	1979	1980	1985
Frasch	6.4	6.8	6.8
Recovered Elemental	<u>4.1</u>	<u>4.4</u>	<u>6.5</u>
Total Elemental	10.5	11.2	13.3
Sulfur in Other Forms	<u>1.7</u>	<u>1.6</u>	<u>2.0</u>
Total Production	12.2	12.8	15.3
CONSUMPTION (END-USE)			
Sulfuric Acid			
Phosphates	8.5	9.0	9.7
Other	<u>3.7</u>	<u>3.6</u>	<u>3.8</u>
Total	12.2	12.6	13.5
Non-Acid	<u>1.1</u>	<u>1.0</u>	<u>1.5</u>
Total	13.3	13.6	15.0

TABLE 4-4. RECOVERED ELEMENTAL SULFUR IN THE UNITED STATES SHIPMENTS, VALUE OF SHIPMENTS, AND AVERAGE VALUE PER TON BY STATES, 1976 AND 1977 (QUANTITIES IN THOUSANDS OF LONG TONS, TOTAL VALUE IN THOUSANDS OF DOLLARS)

STATE	1976				1977		
	SHIPMENTS	VALUE FOB PLANT			SHIPMENTS	VALUE FOB PLANT	
	QUANTITY	TOTAL VALUE	AVERAGE VALUE PER TON		QUANTITY	TOTAL VALUE	AVERAGE VALUE PER TON
ALABAMA	206	9 441	45 83		288	12 761	44 58
CALIFORNIA	432	7 940	18 38		499	9 398	18 83
ILLINOIS AND INDIANA	223	7 470	33 49		242	8 469	34 99
KANSAS	7	253	36 14		5	157	31 40
LOUISIANA	122	6 228	51 05		157	7 281	46 38
MICHIGAN AND MINN	58	1 867	32 19		75	2 375	31 67
MISSISSIPPI	224	11 264	50 29		335	16 655	49 72
NEW JERSEY	188	5 043	46 69		126	5 920	46 98
NEW MEXICO	45	1 480	32 89		57	1 782	31 26
OHIO	17	714	42 00		22	965	43 86
OKLAHOMA	9	324	36 00		10	364	36 40
PENNSYLVANIA	91	3 876	42 59		81	3 467	42 80
TEXAS	872	34 297	39 45		982	38 709	39 42
OTHER STATES ¹⁾	733	28 024	38 23		696	25 545	36 70
TOTAL	3 147	118 221	37 57		3 567	133 848	37 52

1) COMBINED TO AVOID DISCLOSING INDIVIDUAL COMPANY CONFIDENTIAL DATA. INCLUDES ARKANSAS, COLORADO (1977), FLORIDA, DELAWARE, KENTUCKY (1977), MISSOURI, MONTANA, NEW YORK, NORTH DAKOTA, UTAH, VIRGINIA, WASHINGTON, WISCONSIN, WYOMING, VIRGIN ISLANDS, AND PUERTO RICO

SOURCE: U.S. BUREAU OF MINES, SULFUR IN 1977 MINERAL INDUSTRY SURVEYS, MAY 1, 1978

Sulfur in all forms was produced in 1977 by 74 companies at 190 operations in 33 states, according to the U.S. Bureau of Mines. The ten largest producing companies, with 55 of these operations, represented about 75 percent of total industry output. The states of Louisiana and Texas accounted for 67 percent of total production.

Elemental sulfur, often termed crude sulfur or brimstone, presently accounts for 89 percent of total output of sulfur. The balance is represented by production of non-elemental or non-brimstone sulfur contained mostly in by-product sulfuric acid from metal smelters and also from pyrites, hydrogen sulfide and liquid sulfur dioxide.

Alabama and Mississippi are fairly large producers of recovered elemental sulfur as shown on Table 4-4. Also the value of sulfur at the plant is significantly larger for Alabama and Mississippi than other states.

As was indicated, most sulfur consumption is in the form of sulfuric acid. The end uses of elemental sulfur (1977) sold or used in the U.S. (Reference 5) are shown in Table 4-5.

TABLE 4-5. SULFUR END USES

	THOUSAND LONG TONS	PERCENT OF TOTAL
Sulfuric Acid	9,446	85.0
Agricultural Chemicals	253	2.2
Paints, Explosives, Organic Chemicals, etc.	271	2.3
Other Industrial Inorganic Chemicals	217	1.9
Pulp and Paper Products	145	1.3
Petroleum Refining, Petro- leum and Coal Products	109	0.9
Synthetic Rubber, Cellulosic and Plastic Products	64	0.5
Food and Kindred Products	5	---
Unidentified	667	5.9
TOTAL	11,117	100.00

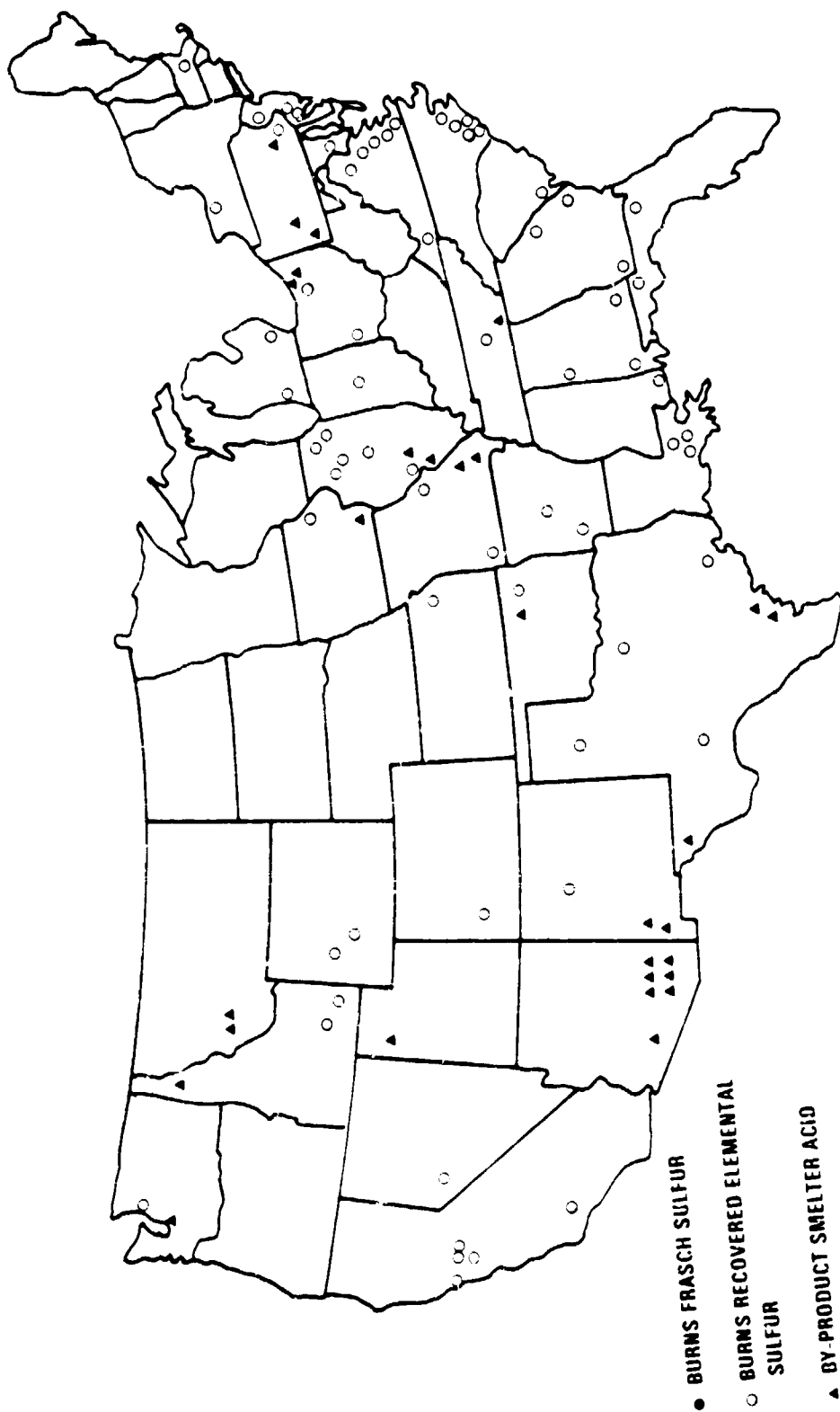
The production of sulfuric acid in the United States has increased at an average annual rate of about 2.5 percent over the last 10 years although this growth has been slightly erratic in certain years.

Derivative phosphoric acid production is currently running at near optimum capacity (Reference 6). This requires high capacity use of feeder sulfuric acid plants. Behind the rising capacity use at sulfuric acid plants are two forces, booming exports in phosphate fertilizations and swollen demand from U.S. manufacturing. Both influences are cyclical, even volatile. Downswings could affect sulfuric acid production quickly. In the U.S. and export markets, so much sulfuric acid demand hinges on fertilizer use, roughly two-thirds of the total, that other uses seldom have much effect on the overall market. With the lone exception of uranium ore processing at times, sulfuric acid's growth is tied firmly to phosphoric acid used for fertilizer.

Sulfuric acid and by-product smelter acid producers are shown on Figure 4-2. A large by-product smelter acid facility is located at Copperhill, Tennessee, producing about 1,250,000 tons per year. The four sulfur-burning sulfuric acid plants in Alabama produce 342,000 tons per year. The one sulfur-burning Tennessee plant produces 132,000 tons per year (Reference 7).

By far the largest end-use of sulfuric acid (61 percent of the total) is in the manufacture of inorganic fertilizers and intermediates such as phosphoric acid and phosphate fertilizers. Industrial users of sulfuric acid are extensive and are concentrated largely in the major industrial regions of the country, but no single end-use represents more than 7 percent of total demand. Most industrial uses are well developed, and none have projected growth rates which are expected to reverse the trend of increasing percentages of sulfuric acid being consumed for fertilizers. A more concise summary of the end-use pattern for 1977 (from Bureau of Mines) is provided in Table 4-6.

About 90 percent of the sulfuric acid used in fertilizers is in the manufacture of wet process phosphoric acid by the reaction of sulfuric acid with phosphate rock. Also in the fertilizer category, but much smaller in volume, is the production of ammonium sulfate by the reaction of sulfuric acid with ammonia.



SOURCE: TENNESSEE VALLEY AUTHORITY

Figure 4-2. Locations of Sulfuric Acid Plants and By-Product Smelter Acid Producers

TABLE 4-6. ESTIMATED 1977 END USES OF SULFURIC ACID

DERIVATIVE	PERCENT OF TOTAL
Fertilizers	61
Alkylation (Petroleum Products)	7
Copper and Other Nonferrous Metals	6
Paints and Pigments	2
Pulp and Paper	1
Steel Production	1
Plastic and Synthetic Products	1
Other Chemicals	8
Other Miscellaneous Uses	<u>12</u>
TOTAL	100

Oxygen

Oxygen is the largest volume U.S. industrial gas, followed closely by nitrogen. Oxygen usage has been growing steadily for some time. In 1975, oxygen usage was down (Reference 8) because of the economic recession. Also, oxygen usage follows the steel industry, and steel production is falling off. However, oxygen sales have, in general, remained profitable. Most oxygen sold to the steel industry is on-site or tonnage oxygen. Tonnage oxygen is produced by an air separation plant at the steel site and is simply placed into the blast furnace. The oxygen from these "dedicated" plants is sold to customers under take-or-pay contracts. The only time overall oxygen profits are affected during an economic slow-down is at the beginning, when steel companies cut their orders for merchant oxygen supplementing tonnage oxygen at their plants. Merchant or open-market oxygen is brought in by truck. Merchant oxygen is expected to grow 7% to about 68 billion cubic feet a year (Reference 8).

Merchant oxygen prices are currently 33.5 to 34 cents per hundred cubic feet (plus transportation and equipment charges) for the largest users. Significant new oxygen capacity is coming on line for merchant marketing. There are many uncertainties in future oxygen demand; a recession, steel production cutbacks, higher electrical costs. All can drive oxygen demand down.

Primary metals industry utilize 30% of oxygen production, followed by health services (13%), and metal fabricating (6%).

Nitrogen

Nitrogen demand growth has been outstanding since the early 1970's. Average production growth has been averaging 11% or greater per year. Nitrogen's growth has been so good it is expected to overtake oxygen as the largest industrial gas (Reference 8). As with oxygen, merchant production of nitrogen is most important to producers, since tonnage nitrogen again is sold on commodity-type, lower profit take-or-pay contracts. Recently, however, the merchant market for nitrogen has been lagging the total market.

The growth of nitrogen demand, especially tonnage nitrogen, has been tied to the increasing acceptance as an inert blanketing agent for metals, electronics, and chemical processing. Nitrogen has been replacing natural gas as a blanketing agent, and now nitrogen's biggest market is as a blanketing agent. Another fast growing nitrogen market is as a freezing agent, especially in processed foods.

Price increases for nitrogen have been lagging general inflation because of production economies of scale. Current prices (mid-1979) for merchant nitrogen are 30.5¢ per hundred cubic feet, plus transportation and equipment charges. The price for tonnage nitrogen is much lower and is specified in a long-term contract.

Plant capacity utilization has little to do with the nitrogen price, except in a few isolated cases. Generally, air separation plants can produce more than three times more nitrogen than oxygen. Therefore, producers of nitrogen are in a position in which their product generally exceeds the existing market.

Carbon Dioxide

Carbon dioxide is an important industrial gas, usually generated from natural sources and recovered in small amounts for commercial uses. Most carbon dioxide used in merchant or captive consumption is obtained as a by-product of the production of synthesis gas which is used for producing ammonia, methanol and other chemicals, and also for hydrocracking and hydrotreating of petroleum products. Large amounts of industry-generated carbon dioxide are vented to the atmosphere. Carbon dioxide is not manufactured by merchant producers, but is gathered by them as a by-product from the basic processes of other companies. Therefore, the cost of merchant carbon dioxide is actually based on the liquefaction capacity required, since the raw gas is taken from ammonia plants and synthetic gas facilities. Carbon dioxide plants tend to be situated near markets because it is a cheap product and transportation costs are relatively high. In general, it is not economical to ship carbon dioxide more than 200 to 300 miles. Plants vary in capacity from a few tons per day up to about 700 tons per day (Reference 9). A summary of plants and production amounts for regions in the U.S. during 1977 are shown in Table 4-7.

TABLE 4-7. REGIONAL PATTERN OF U.S. PRODUCTION AND SHIPMENTS OF
LIQUID AND SOLID CARBON DIOXIDE IN 1977

REGION AND STATES	NUMBER OF PLANTS	PRODUCTION (THOUSAND TONS)	SHIPMENTS (THOUSAND TONS)	VALUE OF SHIPMENTS (MILLION DOLLARS)
NEW ENGLAND & MIDDLE ATLANTIC (ME, NH, VT, MA, RI, NY, NJ, PA, CT)	22	1136	1079	106
E NORTH CENTRAL (OH, IN, IL, MI, WI)	36	4123	3975	183
W NORTH CENTRAL (MN, IA, MO, ND, SD, NB, KS)	13	2749	2683	113
S ATLANTIC & E SOUTH CENTRAL (DE, MD, DC, VA, W VA, NC, SC, GA, FL, TN, AL, MS, FL, KY)	40	5237	4927	234
W SOUTH CENTRAL (AR, LA, OK, TX)	36	5975	4588	164
MOUNTAIN (MT, ID, WY, CO, AZ, UT, NV)	20	1079	1079	25
PACIFIC (WA, OR, CA, AK, HI)	<u>22</u>	<u>2259</u>	<u>2170</u>	<u>58</u>
TOTAL U.S.	195	22558	20901	883

SOURCE: U.S. DEPARTMENT OF COMMERCE, BUREAU OF THE CENSUS, CURRENT INDUSTRIAL REPORTS,
SERIES M28C (77)-14

Current uses (Reference 8) for carbon dioxide are for food refrigeration (37%), industrial (15%), beverage carbonation (25%), and other industrial uses (22%). Carbon dioxide is also used in coal mining for pressurizing, for oil well stimulation, and for water treatment.

Generally, purity requirements for carbon dioxide are high. The product usually has to be 99.98% pure carbon dioxide with all sulfur removed since trace sulfur is unacceptable for carbonated beverages, food processing, and refrigeration uses.

Patterns of merchant consumption for carbon dioxide use is shown in Table 4-8. The growth of the fast food industry has boosted carbon dioxide usage in refrigeration and carbonation. The merchant market for carbon dioxide is less than half of the total. Area production, chemical industries, and oil well stimulation account for about half of the total market.

Steam

Most industries use hot water, steam, or hot air in space heating, process heating, or as a process feedstock. The nationwide utilization of industrial process heat is shown on Table 4-9 (Reference 10). Steam usage is projected to rise to about 22 percent of the total process heat utilization by 1985. The big users of steam are the chemical, petroleum, and textiles industries.

The value of steam is based on its utility, which is a function of both its enthalpy and end-use. Most plants use several steam-pressure levels, and in many plants, waste-heat boilers may supply steam at several different levels. Placing a market price on steam is rather difficult. Process steam from waste-heat recovery may have a low effective generation cost because the heat would otherwise be rejected to the surroundings. Process steam from fossil or synthetic-fuel fired boilers will have a much higher cost because of the high cost of the imported fuel. The average cost of producing steam in a typical plant is estimated to be \$5-6 per MMBTU in 1980.

Because steam must be transported through insulated piping to conserve its energy, industries which might purchase the steam will have to be collocated with the coal gasification facility. An industrial reserve site is in fact located southwest of the planned Murphy Hill site.

Tars

Coal tars are usually made from the higher-boiling fractions (over 450°F) of the crude tar produced by coke ovens and coal retorts. The available grades range from liquids at 30 sec. viscosity at 100°F to a pulverized pitch. Chemically, they are mixtures of hydrocarbons, phenols, and heterocyclic nitrogen, oxygen, and sulfur compounds. They are much more aromatic than petroleum fuels and burn with a luminous flame. Sulfur and ash contents are usually low. The ash, unlike petroleum ash, is high

TABLE 4-8. TRENDS IN MERCHANT CARBON DIOXIDE CONSUMPTION
(PERCENTAGES OF TOTAL BY MAJOR END USES)

	1964	1967	1973	1977	1978
<u>REFRIGERATION</u>					
FOOD	39	40	30	30	30
INDUSTRIAL	1	1	10	10	15
TOTAL REFRIG.	39	40	40	40	45
<u>CARBONATION</u>	20	20	20	35	35
TOTAL MAJOR USES	59	60	60	75	80
<u>OTHER USES</u>					
CHEMICAL RAW MATERIAL	14	14	10	1)	1)
INERTING 2)	10	10	10	1)	1)
PRESSURIZING	5	5	5	1)	1)
OIL WELL STIMULATION	—	—		1)	1)
MISCELLANEOUS	12	11	15	1)	1)
TOTAL OTHER USES	41	40	40	25	20
TOTAL OTHER USES	100	100	100	100	100

1) BREAKDOWN OF "OTHER USES" NOT AVAILABLE

2) INERT ATMOSPHERES FOR FOOD PROCESSING, FIRE EXTINGUISHING, ETC

SOURCES CHEMICAL ECONOMICS HANDBOOK 1974 p 731 5010 3.
CHEMICAL & ENGINEERING NEWS JUNE 26 '78 p 13, JULY 18 1977, p 11

TABLE 4-9. INDUSTRY PROCESS HEAT UTILIZATION (PROJECTED TO 1985, 10^{12} BTU/YEAR)

INDUSTRY/SEGMENT	HOT WATER 212 F	STEAM		DIRECT HEAT/HOT AIR			ROUNDED TOTALS
		212-350 F	350 F	212 F	212-350	350 F	
ALUMINUM			75			125	200
AUTOMOBILES AND TRUCKS	15.1	1.6		24.7	11.6	1.0	54
CEMENT						490	490
CERAMIC						370	370
CONCRETE BLOCK AND BRICK		17	8				25
GYPSUM			38		17		55
CHEMICALS (INORGANIC)		286		5.5	14.5	234	540
COAL MINING AND CLEANING						18	18
COPPER	14.2			2.4		55.4	72
FOOD PROCESSING	22	98	36.4	4.0	34	5.6	200
GLASS	12	14			14	446	486
LUMBER	4.3	16.2	3.1	83.1	3.2	55.1	165
MINING (FRASCH SULFUR)		51					51
PAPER AND PULP		645				130	775
PETROLEUM REFINING		150	450			3200	3800
PLASTICS/SELECTED POLYMERS		9.4	11.6				21
RUBBER/SBR MANUFACTURE		8.3					8
STEEL AND IRON		75				1925	2000
TEXTILES	25.7	258	5.8		93.6	16.9	400
ROUNDED TOTALS	93	1630	628	120	187	7072	9,730
PERCENT OF TOTAL	1.0	16.8	6.5	1.2	1.9	72.6	

A-4-18

6-2

in alkaline earth elements and usually full of vanadium. Typical analyses (Reference 11, page 9-11) are shown on Table 4-10.

Because of natural gas curtailments and fuel costs, many industries are driven to using No. 6 fuel oil. Fuel oil usage in northern Alabama and south central Tennessee in 1977 was approximately 20 trillion BTU/year. If the coal tar can be converted to a fuel oil, a significant market exists. Table 4-10 illustrates a rough comparison between Lurgi-derived coal tars and number 6 fuel oil.

Argon

Argon is an abundant member of the rare gas family, and is a colorless, tasteless gas somewhat soluble in water. Argon is typically sold in four commercial grades depending on purity, nitrogen, oxygen, hydrogen, and carbon dioxide content (Reference 12).

Argon is used extensively in the incandescent lamp industry for the filling of light bulbs. It is used in arc welding as an inert gas shield to prevent oxidation of the metals being welded. It is also used with other rare gases in the filling of special bulbs and display tubes to obtain special color effects in the neon-type bulbs. Many Geiger-counting tubes contain argon or argon mixed with organic vapors and gases, particularly, 90% argon-10% methane which is used universally as a proportional counting gas.

Plasma jet torches, utilizing an argon-hydrogen mixture heated to temperatures in excess of 10,000°K, are used for cutting operations and for containing metals with refractory materials. The high-temperature preparation, refining, and fabrication of many materials must be carried out in an argon (or helium) atmosphere. Most of the high-purity single crystals used for semiconducting devices are grown in an argon (or helium) atmosphere. In doping semiconductors with controlled amounts of impurities, the latter are frequently introduced in a stream of argon (or helium). Argon (or helium) is used extensively to carry a reactant to a reaction zone, to modify the rate of a reaction by dilution, and to remove reaction products.

TABLE 4-10. CHARACTERISTICS OF COAL TARS AND NO.6 FUEL OIL

COAL	LURGI-DEVICE COAL-TARS (REF. 12) FUEL OIL				
	MONTANA SUB BITIMINOUS	ILLINOIS #6	ILLINOIS #5	PITTSBURGH #8	NO. 6
DENSITY, LB/U.S. GAL., 60° F	8.8	8.5	9.8	10.3	
WATER CONTENT, WT %	30.0	26.7	10.4	11.9	
ULTIMATE ANALYSIS, %:				88.51	
CARBON	83.06	85.48	85.85	85.51	84.7
HYDROGEN	7.69	6.44	6.40	5.93	11.02
OXYGEN	8.23	5.17	4.16	3.16	0.38
NITROGEN65	1.18	1.19	.87	0.18
SULFUR28	1.70	2.39	1.52	3.97
CHLORINE04	UNDETERMINED	UNDETERMINED	UNDETERMINED	
ASH05	.03	.01	.01	.02
C/H RATIO	10.8	13.3	13.4	14.9	7.62
GROSS CALORIFIC VALUE, B.T.U./LB	15,830	15,900	15,900	16,120	16,000

REFERENCES

1. "Prospective Regional Markets for Coal Conversion Plant Products Projected to 1980 and 1985," Volume 1, Market Analysis, Foster Associates, Inc., November 1974.
2. "Ash - A Thorny Problem in Coal Utilization," Chemical Engineering, January 28, 1980.
3. "Ash/FGD Waste Disposal Options," Combustion, January 1980.
4. "Sulfur: Tight Supplies May Turn Into a Glut," Chemical Engineering, March 24, 1980.
5. "End-Uses of Sulfur and Sulfuric Acid in 1977," U.S. Bureau of Mines, June 20, 1978.
6. "Sulfuric Acid," Chemical and Engineering News, April 30, 1979.
7. "Potential Abatement Production and Marketing of By-Product Sulfuric Acid in the U.S.," TVA, April 1978.
8. "Industrial Gases May Take Slowdown in Stride," Chemical and Engineering News, July 16, 1979.
9. Current Industrial Reports, U.S. Bureau of Census, Series M28C(77)-14.
10. "Industrial Process Heat Survey," Battelle, January 1977.
11. CHEMICAL ENGINEERS HANDBOOK, Fifth Edition, pg. 25-28.
12. "Environmental Assessment Report: High-BTU Gasification Technology," SYMPOSIUM PROCEEDINGS, ENVIRONMENTAL ASPECTS OF FUEL CONVERSION TECHNOLOGY, IV, April 1979, Hollywood, Florida, Report No. EPA-600/7-79-217, September 1979.

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5.0 ALTERNATE PRODUCT ANALYSIS

5.0 ALTERNATE PRODUCT ANALYSIS - TASK 5.1.4

Complete facility economic analysis requires the investigation of possible alternate product options over the production of MBG. Methane, methanol, hydrogen, and gasoline have been identified as possible candidates and have been characterized as in Section 2.0 of this report.

5.1 Approach

The approach taken in characterizing the alternate products options was the same as was used in the MBG case. The primary difference was that the references available to obtain information complete for this section were far less. Appendix A of this report contains an identification of references used in this subtask.

5.2 System Identification

NASA-MARSHALL has identified alternate products as system 20. Table 5-1 is the breakdown that has been used to obtain cost and characterization data for the alternate product options.

5.3 System Description

A brief description of each unit operation has been included in this section. Each description addresses the cost and design drivers and any critical technology issues identified thus far in the study.

5.4 System Characterization

The alternate product unit operations characterizations are reported on the same basis as the MBG unit operations of Task 5.1.1. Appendix E summarizes the stream characterizations and raw materials, and Appendix F the system characterizations. The operation and maintenance costs are reported in Appendix D. Table 5-2 is a summary of the component characterization of the alternate products unit operations.

TABLE 5-1. ALTERNATE PRODUCTS
UNIT OPERATION CATEGORIES

<u>NUMBER</u>	
90	SHIFT CONVERSION
91	METHANATION
92	GAS DRYING
93	METHANOL SYNTHESIS
94	GASOLINE SYNTHESIS
95	HYDROGEN RECOVERY

TABLE 5-2. ALTERNATE PRODUCTS SYSTEM COMPONENT CHARACTERIZATION

	FIELD FABRICATED VESSELS	SHOP FABRICATED VESSELS	COMPRESSORS BLOWERS	RECHARGERS	PUMPS	FIELD EQUIPMENT	SOLID STORAGE	SOLID TRANSFER	PERMEATION	CLIMATE CONTROL	SPECIFICITY (EQUIPMENT)	SIZE REDUCTION CLASS(S)	WASTEAGE
90 SHORT CONVERSION	1 2	4	0	4	B	4	0	0	0	0	0	0	0
91 DETRIMENTATION	1 2	4	2 3	4	C	4	0	0	0	0	0	0	0
92 GAS DRYING	0	4	0	4	4	0	0	0	0	0	4	0	0
93 DETRIMENTATION SYSTEMS	1 2	1 2	2	4	4	0	0	0	0	0	0	0	4
94 GASOLINE SYSTEMS	1 2	1 2	4	4	4	0	0	0	0	0	0	0	4
95 HYDROGEN RECOVERY	0	4	0	4	0	4	0	0	0	0	0	0	0

CHARACTERIZATION CODES: 0 NOT INCLUDED IN SYSTEM
 1 HIGHER COST ITEM
 2 LONG LEAD TIME ITEM
 3 CRITICAL TECHNOLOGY AREA
 4 MORE CRITICAL ITEM

SYSTEM NUMBER 20
UNIT OPERATION NUMBER 20
DESCRIPTION SHIFT CONVERSION

The purpose of this unit is to adjust the hydrogen to carbon monoxide ratio of the raw gas from Gasification for downstream processing. This adjustment is accomplished catalytically by the exothermic "Shift Reaction":



A number of catalysts are commercially available to promote this reaction. Among them are the following:

- Cobalt-molybdate on alumina, high-temperature, 650°F, requires H_2S in feed.
- Iron-chrome, high-temperature, partially deactivated by sulfur in feed.
- Copper-Zinc, low-temperature, cannot tolerate sulfur components in feed.

These tolerances will dictate the placement of the Shift Conversion System relative to the Acid Gas Removal System. All of these catalysts are used in fixed-bed reactors. The process description which follows is for a typical shift conversion system employing a high-temperature cobalt-molybdate catalyst.

Feed gas to the Shift Conversion Unit is preheated by heat exchange against converter effluent gas and the injection of high pressure process steam, to a temperature of about 650°F.

Part of the gas is not preheated totally but bypasses the converters to provide control of the total H_2/CO ratio in the gas leaving the system.

The preheated gas passes through the shift converters, where the shift reaction takes place in a fixed bed of catalyst. The heat of reaction causes the temperature of the outlet gas to rise to about 900°F. This hot gas is then used to preheat the inlet gas and to heat other process streams. At an appropriate point in the gas cooling sequence, the bypassed gas is combined with the shifted gas.

The combined gas stream is further cooled against air or cooling water. Condensed water is separated from the vapor and sent to the water treating systems in the main portion of the plant. The cooled gas then passes to downstream processing.

Depending on the catalyst selected, COS hydrolysis to H_2S may occur. This can reduce the cost of some downstream Acid Gas Removal processes if essentially complete removal of sulfur is required. The bypass gas may also be passed through catalytic hydrolysis for COS conversion to H_2S .

The Design Drivers include:

- 1) H_2/CO ratio requirements in shifted gas
- 2) Feed Gas Temperature, °F
- 3) COS hydrolysis requirements
- 4) Catalyst selection.

The Cost Drivers include:

- 1) CO moles shifted
- 2) COS moles shifted in bypass.

Critical Technology Issues which must be addressed are the amount of COS hydrolysis to be expected in commercially available cobalt-molybdate shift catalysts, and the commercial availability of COS hydrolysis catalysts for use on the bypass stream.

SYSTEM NUMBER 20
UNIT OPERATION NUMBER .91
DESCRIPTION

METHANATION

The purpose of the Methanation Section is to effect the requisite chemical reactions for converting the purified synthesis gas into a high-Btu, "pipeline quality" gas which is essentially equivalent to natural gas. Nickel catalyst and temperature-controlled, fixed-bed reactors are used to effect the reactions.

The essence of any methanation process is to control the heat of reaction so that the process operates at the desired temperatures and to recover the exothermic heat of reaction so that a high thermal efficiency is realized. The methanation process is a catalytic fixed-bed, adiabatic, gas-recycle process. A highly active nickel catalyst is used to effect the methanation reactions. Reaction temperatures are controlled by recycling a portion of cooled product gas.

The process consists of three major subsystems:

- Methanation reactors
- Gas recycle to methanation reactors
- Waste heat recovery.

A large number of possible configurations are possible. Some of these involve proprietary technology, others do not. The following process description is based on a non-proprietary configuration using a commercially available, but as yet unproven, catalyst. To date, no methanation catalyst has been commercially demonstrated for high-BTU pipeline gas production from coal-derived synthesis gas.

Feed gas is preheated by hot process gas and passes through sulfur guard reactors which remove traces of sulfur from the gas. This is required to avoid poisoning the methanation catalyst. The desulfurized feed gas passes through a series of fixed-bed catalytic reactors, combining with portions of fresh feed and recycle gas between reactors. The methanated effluent gas is cooled, compressed, and split into product and recycle gas streams.

After each methanation reactor, heat of reaction is removed from the effluent by generating high-pressure steam in a shell-and-tube exchanger. After being split from the recycle gas, the net product is cooled by preheating the guard bed feed, then by cooling with water or air. Condensed water is separated from the gas stream, which may then be compressed to pipeline pressure.

Design drivers for the unit are:

- 1) Type of feed/recycle splitting
- 2) Catalyst selection and properties
- 3) H_2/CO ratio in feed
- 4) Feed Pressure
- 5) Utility of HP steam in the plant.

Cost drivers are:

- 1) Feed rate, ACFH
- 2) Feed pressure
- 3) Moles CH_4 produced
- 4) Feed sulfur content.

A critical technology issue concerns the commercial availability of a high-temperature recycle compressor. If such equipment were not available, cold gas recycle would be used and a significant energy penalty would result from the loss of latent heat in the water condensed from the recycle gas.

SYSTEM NUMBER 20
UNIT OPERATION NUMBER 92
DESCRIPTION PRODUCT GAS DRYING

The purpose of this unit is to dehydrate the SNG product to the pipeline specification using TEG (triethylene glycol).

This is a non-proprietary process.

The product gas is passed countercurrent against a TEG solution (lean solution) in a TEG absorber giving up water to the TEG solution. The dry gas is routed to the pipeline.

The water-laden TEG solution (rich solution) leaves the TEG absorber after being countercurrently contacted with the product gas stream. The rich solution is depressurized and heated by coils in the TEG Storage Drum and the TEG Regenerator. In the TEG Storage Drum, the rich glycol is heated up by the hot, lean TEG in the storage drum. The coil in the regenerator serves a dual purpose. It heats up the rich glycol solution before it enters the regenerator, and it also serves as an overhead condenser for the regenerator. The water vapor from the regenerator is vented to the atmosphere.

The TEG regenerator is reboiled with 550 psig steam. Enough steam is used to achieve proper concentration of the glycol solution.

The lean glycol solution leaves the reboiler and is cooled by the rich glycol circulating in the coil of the storage drum. The lean solution is then pumped above pipeline grid pressure to enter the top of the absorber. A reciprocating pump is used for this purpose. The lean solution re-enters the absorber after being cooled by a coil in the absorber near the vapor outlet thus completing the TEG circuit.

The primary cost and design drivers are quantity of gas to be dried and the water content of the product gas.

SYSTEM NUMBER 20
UNIT OPERATION NUMBER 93
DESCRIPTION METHANOL SYNTHESIS

The purpose of this unit is to convert purified synthesis gas into methanol using a catalytic reaction.

A number of proprietary processes are commercially available from several licensors, including Lurgi and Davy-McKee, Inc. The following description is for the Lurgi 50-atmosphere synthesis process, which has been commercially demonstrated.

The synthesis gas entering the system is compressed to about 1100 psig and mixed with a recycle gas stream. A recycle gas compressor then compresses the mixture to overcome the pressure losses of the recycle loop. This loop consists of a heat exchanger to heat syngas and recycle gas against methanol and recycle gas, a fixed-bed catalytic reactor, condensers, and a separator in which the unconverted gas is separated from the methanol for recycling.

In the reactor, the heat of reaction is transferred to boiler feed water flowing outside the catalyst tubes, thereby generating medium pressure steam. The reactor temperature is controlled by controlling the steam pressure.

The high pressure methanol liquid product is reduced in pressure and flows into a vessel for degassing; the methanol is sent to By-Product Storage, Unit Operation 82.

To eliminate the build-up of methane and inerts in the recycle gas, a portion of the recycle gas is purged continuously and used for plant fuel along with the gas from the methanol separation vessel.

The primary cost and design drivers are:

- Capacity, TPD Methanol produced
- Feed gas pressure
- Percent inerts (compounds other than H_2 and CO) in the feed gas.

SYSTEM NUMBER 20
UNIT OPERATION NUMBER 94
DESCRIPTION GASOLINE SYNTHESIS (FROM METHANOL)

The purpose of this unit is to dehydrate crude methanol to LPG and gasoline hydrocarbons which can be marketed directly without further processing.

Mobil has demonstrated a proprietary catalytic process to dehydrate methanol to the stoichiometric yield of hydrocarbons (44%) and water (56%). The reaction takes place over fixed-bed catalytic reactors. The reaction is highly exothermic with the heat of reaction being about 740 Btu/lb of methanol. Therefore, the principal problem in designing a reactor system is heat removal. In the fixed bed reactor system, this problem is minimized by dividing the above overall reaction (and reactor system) into two steps. In the first, methanol is partially dehydrogenated to form an equilibrium mixture of methanol, dimethyl ether and water. About 20% of the heat of reaction is released in this first step without any recycle for temperature control. In the second step, where hydrocarbon conversion takes place, a large recycle gas stream (from 6/1 to 9/1 ratio) is used to limit the temperature rise across the bed to about 125°F.

While other systems, such as quenched reactor, tubular heat exchange reactor, or fluidized bed reactor could be designed, the fixed bed, gas recycle system has been well demonstrated and is readily scaleable to higher throughputs.

Small amounts of CO, CO₂ and coke and trace amounts of formic acid and acetone in the water product are formed as by-products. After fourteen days on stream, enough coke accumulates on the catalyst to require regeneration. The catalyst is regenerated by coke combustion in a controlled manner to limit the maximum temperature to 900°F. Since the catalyst converts other oxygenates and is tolerant to water, crude methanol without any purification can be charged directly, thereby simplifying the methanol synthesis.

The produced hydrocarbons are predominantly in the gasoline boiling range (C₄ to C₁₀), and the gasoline is chemically conventional, consisting of highly branched paraffins (51%), highly branched olefins (13%), naphthenes (8%) and aromatics (28%). Its unleaded research octane number (RON) ranges from 90 to 100. Essentially, no hydrocarbons larger than C₁₀ and no oxygenates are produced.

Other minor subsystems are used to fractionate the hydrocarbon product and to catalytically upgrade the fractionated products to export quality.

The major cost and design driver for this system is the methanol feed rate, tons per day.

SYSTEM NUMBER 20
UNIT OPERATION NUMBER 95
DESCRIPTION

HYDROGEN RECOVERY

The purpose of this unit is to separate hydrogen from other gases, to produce the appropriate purity hydrogen for gas sales.

The Pressure Swing Absorption Process (PSA) is commercially available from the Union Carbide Corporation.

The main equipment elements of this unit are parallel pressure-swing absorption columns which alternate absorption and regeneration. During operation, essentially all non-hydrogen elements and a portion of the hydrogen are absorbed. The remaining hydrogen leaves the unit at a purity of 98+%.

Regeneration is by pressure reduction in 3 stages with the final pressure nearly atmospheric. The desorbed gases can be compressed and recycled to an upstream shift conversion unit, or used for plant fuel.

The primary cost and design drivers are:

- % H₂ in feed gas and purity of product hydrogen
- Capacity of the unit, gas throughput
- Pressure and temperature of feed.

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6.0 PRELIMINARY INTEGRATED FACILITY REQUIREMENTS DOCUMENT

6.0 PRELIMINARY INTEGRATED FACILITY REQUIREMENTS DOCUMENT
(PIFRD) - TASK 5.1.5

This document has been prepared in preliminary form using a draft provided by NASA-MARSHALL and completed using industry experience and reference literature. The two PIFRD's have been based on Koppers-Totzek and Texaco gasifier technology and have been submitted, in DR-11 format, separate from this document.

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APPENDICES

APPENDIX A REFERENCES

Ref. No.	Report No.	Gasifier Type	Product Gas	Plant Type	Coal	Date	A/E	Remarks
1.	EPRI AF-741	Koppers-Totzek	HBC	Liquefaction	Illinois #6	76	Parsons	
2.	EPRI AF-416	Lurgi, BGC Slagging, Foster Wheeler Entrained Flow		Combined Cycle	Illinois #6	75	Stone & Webster	
3.	7E 2447-13	Lurgi		Mobil Methanol to Gasoline & Fischer-Tropsch to produce gasoline, SMG, LPG	Sub-bituminous Wyoming	77	Mobil	
4.	PS 264 702	L, K-T, S&W, Winkler, Weillman-Golusha	LBC HBC	LBC, IMG Production for iron ore pelletizing	Lignite, Sub-bituminous, Bituminous	76	McKee	Limited cost information on K-T only
5.	EPRI AF-442	L, BGC Slagger, Foster-Wheeler, CE, T		Combined Cycle	Illinois #6, New Mexico	76	Fluor	
6.	EPRI AF-244	L, IGT, CE		Fuel Gas Production	Illinois #6	75	Fluor	
7.	EPRI AF-511	Lurgi, Koppers-Totzek	HBC	Retrofitting Power Plant	1.5% S coal	75	TVA	
8.	FE 1775-13	2-stage entrainment slagger		POGO (Power-Oil-Gas-Other) Design Facility	1) Eastern Interior Province 2) S. Appalachian Eastern Province 3) Powder River Rocky Mt. Province	77	Parsons	
9.	FE 1775-18	Modified Bi-Gas IGT Design	LBC, HBC SMC	MPGP (Multi-Process Demonstration Plant) gasification, combined-cycle, Synpas production, Fischer-Tropsch Indirect Liquefaction	Eastern Interior Province	77	Parsons	
10.	EPRI AF-916	Tenaco	HBC	Combined Cyc's	Illinois #6	76	Fluor	
11.	EPRI AF-880	Tenaco			Low-sulfur Western	76	Parsons	
12.	FE 1775-7	Bi-Gas		Fisher-Tropsch Oil and SMG Production	Eastern Interior Province	77	Parsons	
13.	FE 2240-31	IGT SYGAS, BCR Bi-Gas, PERC Synthene, Lurgi		SMG Production	Pittsburgh #8	76	Braun	
14.	NEIGH DEMO (Vol. 1)	U-Gas	HBC	Produce Industrial Fuel Gas	High Sulfur Bituminous	79	Foster Wheeler	Limited System Information, No costs given
15.	GRACE DEMO	Tenaco	HBC	Demo Plant		79	ESBSCO	Flow sheets describe various systems and their components, but no cost information
16.	PROCON CCD	HYGAS	SMC	Demo Plant	Bituminous and Lignite	78	Procon	
17.	COMOCO/BGC CCD FE-2542-10 (Vol. 2)	BGC, Lurgi Slagger	SMC	Demo Plant	Illinois #6 (sensitivity of costs to Ohio 29 and Pittsburgh #8 given)	78	Foster Wheeler	
18.	FE 2542-13					77	Continental Oil	Data base for Pipeline Demo Plant, no cost information
19.	EPRI AF-753	Tenaco	HBC	Combined Cycle	Illinois #6	76	Fluor	
20.	EPRI AF-791					-	Gibbs & Hill	Discusses preparation of coal for combustion and conversion. Gives equipment and cost information for coal preparation system for coals and gasifiers of interest.
21.	EPRI AF-615	Koppers-Totzek	HBC	Char Gasification using COED Process	Char W. Kentucky Pittsburgh	78	Koppers	Specific equipment and cost data not available
22.	EPRI AF-1227	Tenaco	Methanol Fuel Oil		Illinois #6 Wyodak Sub-bituminous	77	Braun	
	EPA-600/7-78-010	None	None	Coal-Fired Boiler With FGD	Eastern 4.8% Sulfur	78	A. J. Little	CEA Dual-Alkali
23.	APPENDIX A REFERENCE CHARACTERIZATION							

ORIGINAL PAGE IS OF POOR QUALITY

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 2
UNIT OPERATION NUMBER 10
DESCRIPTION

COAL RECEIVING, STORAGE AND TRANSFER

SHEET 1 OF 2

REFERENCE/CASE	1/1-4	5/MXSC	13/MIGAS	13/LARGE	6/MX	6/FX	6/EXL
NO. OF PARALLEL TRAINS OPERATING/SPARE					1/0		
SYSTEM COST, MM \$	10.749	14.905	12.00	12.00	12.441	15.645	23.828
DATE OF MONEY	Mid 76	Mid 76	1/76	1/76	Mid 75	Mid 75	Mid 75
COST DRIVERS							
1. MAX CAPACITY, TONS/DAY	19,681		38,400	36,000			
2. STORAGE, MTONS *	760		226.4	255.4			
3.							
4.							
5.							
DESIGN DRIVERS							
1. DELIVERY METHOD	Unitrain bottom						
2.	dump						
3.							
4.							
5.							
SYSTEM INTERFACES							
1. ELECTRICITY, KW	1556						
2. H.P. STEAM, Mlb/HR							
3. M.P. STEAM, Mlb/HR							
4. L.P. STEAM, Mlb/HR (PSIG)							
5. BOILER FEEDWATER, GPM							
6. COOLING WATER, GPM							

* These cost drivers are also design drivers.

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 2

UNIT OPERATION NUMBER 10

DESCRIPTION COAL RECEIVING, STORAGE AND TRANSFER

SHEET 2 OF 2

REFERENCE/CASE	8/	11/1	12/	16/Bituminous			
NO. OF PARALLEL TRAINS OPERATING/SPARE				3/0			
SYSTEM COST, MM \$	36.712	8.780	23.0	96.240			
DATE OF MONEY	Mid 77	Apr 78		1st Qtr 78			
COST DRIVERS	1. MAX CAPACITY, TONS/DAY	43,700	1,000	48,000	21,320		
	2. STORAGE, MTONS *	280	3	0	640		
	3. AVAILABLE AREA, SQ FT						
	4.						
	5.						
DESIGN DRIVERS	1. DELIVERY METHOD		Unitrain 100T Bottom Dump				
	2.						
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	14,000		8,400			
	2. H.P. STEAM, Mlb/HR			0			
	3. M.P. STEAM, Mlb/HR			0			
	4. L.P. STEAM, Mlb/HR(PSIG)	378.5(135)		0			
	5. BOILER FEEDWATER, GPM			0			
	6. COOLING WATER, GPM			1,120 (makeup)	from C-T blowdown		

* These cost drivers are also design drivers

A-B-2

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 11

UNIT OPERATION NUMBER 11

DESCRIPTION COAL PREPARATION AND FEEDING

SHEET 1 OF 1

REFERENCE/CASE	1/1	1/2	1/3	1/4	11/1	12/	13/BIGAS
NO. OF PARALLEL TRAINS OPERATING/SPARE	2/1	2/1	2/1	2/1	2/0	2/0	1/0
SYSTEM COST, MM \$	26.816	26.816	26.816	26.816	4.970	12.2	12.0
DATE OF MONEY	Mid 76	Mid 76	Mid 76	Mid 76	Apr 78	1st Qtr 75	1/76
COST DRIVERS	1. CAPACITY, TONS/DAY	19,681	20,272	19,993	22,783	975	30,000
	2. COAL SOURCE (ROM OR WASHED)						38,400
	3. FINES MINIMIZATION						
	4.						
	5.						
DESIGN DRIVERS	1. FINES MINIMIZATION						
	2. COAL PROPERTIES						
	3. BLENDING REQUIREMENTS						
	4. WASTE HEAT AVAILABILITY						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	1,603	1,603	1,603	1,603	564	8,100
	2. H.P. STEAM, lb/HR						
	3. M.P. STEAM, lb/HR						
	4. L.P. STEAM, lb/HR (PSIG)						
	5. BOILER FEEDWATER, GPM					67	140,000 (50)
	6. COOLING WATER, GPM						

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 12

UNIT OPERATION NUMBER 20

DESCRIPTION GASIFICATION - KOPPERS-TOTZEK

SHEET 1 OF 6

REFERENCE/CASE	1/1**	1/2	1/3	1/4	4/ K-T	4 K-T	4 K-T
NO. OF PARALLEL TRAINS	4/1	4/1	4/1	4/1	4/1	4/1	4/1
OPERATING/SPARE							
SYSTEM COST, MM \$*	77.290	106.750	91.838	136.381	54.5	59.5	54.5
DATE OF MONEY	Mid 76	Mid 76	Mid 76	Mid 76	4th Qtr 76	4th Qtr 76	4th Qtr 76
COST DRIVERS	1. CAPACITY, COAL FEED, TPD	3,552	4,256	3,819	5,431	3,400	3,400
	2. O ₂ FEED, lb/HR	143,333	180,000	148,750	252,500	202,974	195,477
	3. STEAM FEED, lb/HR (PSIG)	-	-	-	-	39,151	75,248
	4. GASIFIER PRESSURE (PSIG)					0.3	0.3
	5.						
DESIGN DRIVERS	1. GASIFIER TEMP, °F	3,500	3,500	3,500	3,500	3,300	3,300
	2. COAL TYPE	ILL. NO. 6	ILL. NO. 6	ILL. NO. 6	ILL. NO. 6	LIGNITE	BITUMINOUS
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	31,983	51,165	42,295	72,071		
	2. H.P. STEAM, Mlb/HR(PSIG)					-303.2 (600)	-352.8 (600)
	3. M.P. STEAM, Mlb/HR(PSIG)	-312.4 (600)	-422.7 (600)	-352.8 (600)	-591.5 (600)		-303.2 (600)
	4. L.P. STEAM, Mlb/HR(PSIG)	9.200 (60)	-79.9 (60)	91.3 (60)	-49.3 (60)		
	5. BOILER FEEDWATER, GPM	6,400	6,440	5,370	9,380	-64.9	-56.7
	6. COOLING WATER, GPM						-48.2

* Includes cooler, solids handling and gasifier.

** Includes compressor also.

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - KOPPERS-TOTZEK

SHEET 2 OF 6

A-B-5	REFERENCE/CASE		7/2						
	NO. OF PARALLEL TRAINS								
	OPERATING/SPARE		11/1						
	SYSTEM COST, MM \$ *		56.680						
	DATE OF MONEY		Mid 75						
	COST DRIVERS	1. CAPACITY, COAL FEED, TPD	8,254						
		2. O ₂ FEED, lb/HR	484,000						
		3. STEAM FEED, lb/HR (PSIG)	213,700 (25)						
		4. GASIFIER PRESSURE, PSIG							
		5.							
	DESIGN DRIVERS	1. GASIFIER TEMP, °F MAX	3,300						
		2.							
		3.							
		4.							
		5.							
	SYSTEM INTERFACES	1. ELECTRICITY, KW	7,585						
		2. H.P. STEAM, Mlb/HR(PSIG)	990.7 (1500)						
		3. M.P. STEAM, Mlb/HR(PSIG)							
		4. L.P. STEAM, Mlb/HR(PSIG)	453.4 (30)						
		5. BOILER FEEDWATER, GPM	323						
		6. COOLING WATER, GPM	410						

* Includes cooler, solids handling, and gasifier.

APPENDIX E
SYSTEM CHARACTERIZATION

SYSTEM NUMBER	12	GASIFICATION - TEXACO										SHEET 3 OF 6
UNIT OPERATION NUMBER	20											
DESCRIPTION		5/EXTC	10/1	10/5	11/1							
REFERENCE/CASE		5/1	4/1	4/1	N/A							
NO. OF PARALLEL TRAINS		24.261	22.566	25.658	18.170							
OPERATING/SPACE		Mid 76	Mid 76	Mid 76	Apr 78							
SYSTEM COST, MM \$												
DATE OF MONEY												
1. CAPACITY, COAL FEED, TPD		10,000	915,520	547,776	80,200							
2. O ₂ FEED, lb/hr *		684,720	1200	1200	600							
3. GASIFIER PRESSURE, PSIG		600										
4.												
5.												
1. GASIFIER TEMP, °F		2600	2600	2600	2600							
2. COAL TYPE		Slurry Feed										
3.												
4.												
5.												
1. ELECTRICITY, KW		NA	NA	NA	1787							
2. H.P. STEAM, Mlb/hr (USIG)		-2,117 (1505)	-2,115.5 (1505)	-2,115.4 (1505)	-221.6 (1200)							
3. M.P. STEAM, Mlb/hr		-147.25 (440)	-	-	-							
4. L.P. STEAM, Mlb/hr		-99.00 (50)	-	-	-							
5. BOILER FLEWATER, GPM		4747	4460	4461	445							
6. COOLING WATER, GPM		221	-	178	790							

* These cost drivers are also design drivers

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER		12	SHEET 4 OF 6						
UNIT OPERATION NUMBER		26							
DESCRIPTION		CAPACITATION - LURGI							
REFERENCE/CASE	2/MXL	2/MXH	2/MXL	2/MXH	6/MX	7/1			
NO. OF PARALLEL TRAINS	NA	NA	NA	NA	4/0	11/1			
OPERATING/STABLE	63.421	54.902	69.932	57.928	66.212	31.932			
SYSTEM COST, MM \$	Mid 75	Mid 75	Mid 75	Mid 75	Mid 75	Mid 75			
DATE OF WORK	14,000	14,000	14,000	14,000	10,000	7,818			
1. CAPACITY, GAL. FLD, TPD	534,000	534,000	534,000	534,000	388,899	213,200			
2. O ₂ FLD, LB/HR *	2,126,400	2,126,400	2,126,400	2,126,400	1,850,700 (340)	711,600 (354)			
3. STEAM FLD, LB/HR DESIGN	NA	NA	NA	NA	NA	300			
4. GASIFIER PRESSURE, PSIG									
5. GASIFIER TEMP, °F	NA	NA	NA	NA	NA	NA			
6. ELECTRICITY, KW									
7. H.P. STEAM, MID/HR									
8. L.P. STEAM, MID/HR (DESIGN)	-450.0 (497)	-450.0 (497)	-450.0 (497)	-450.0 (497)	-323.2 (340)	102.2 (355)			
9. L.P. STEAM, MID/HR	904	904	904	904	772	273			
10. BOILER FEEDWATER, GPM									
11. COOLING WATER, GPM									

* The cost drivers are also design drivers

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - BGC/Lurgi

SHEET 5 OF 6

REFERENCE/CASE NO. (/ PARALLEL TRAINS OPERATING/SPARE SYSTEM COST, MM \$ DATE OF MONEY	2/MSL	2/MSH	2/MSI	2/MSH	5/MXSC	17/	
	NA	NA	NA	NA	3/1	3/1	
	20.837	21.681	25.904	25.519	49.628	49.954	
	Mid 75	Mid 75	Mid 75	Mid 75	Mid 76	1st Qtr 78	
S E R V I C E S	1. CAPACITY, COAL FEED, TPD	15,000	14,000	1,400	1,400	10,000	
	2. O ₂ FEED, lb/HR	534,496	478,272	534,496	478,272	383,934	
	3. STEAM FEED, lb/HR	346,070	309,715	346,070	309,715	248,593	
	4. GASIFIER PRESSURE, PSIG	285	285	285	285	400	635,322 550
	5.						
S E R V I C E S	1. GASIFIER TEMP, °F					3500	
	2.						
	3.						
	4.						
	5.						
S E R V I C E S	1. ELECTRICITY, KW					NA	1,303
	2. H.P. STEAM, lb/HR						
	3. H.P. STEAM, lb/HR(PSIG)	-132,642	-134,596	-132,642	-134,596	-113,000 (400)	
	4. L.P. STEAM, lb/HR(PSIG)						-512,000 (35)
	5. BOILER FEEDWATER, GPM	266	270	266	270	227	1574
	6. COOLING WATER, GPM						13,900

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 12

UNIT OPERATION NUMBER 20

DESCRIPTION

GASIFICATION - BABCOCK & WILCOX

SHEET 6 OF 6

REFERENCE/CASE	4/B&W	4/B&W	4/B&W				
NO. OF PARALLEL TRAINS	Lignite	Bituminous	Sub-bituminous				
OPERATING/SPARE	NA	NA	NA				
SYSTEM COST, MM \$							
DATE OF MONEY							
COST DRIVERS	1. CAPACITY, COAL FEED, TPD						
	2. O ₂ FEED, lb O ₂ /lb COAL	0.80	0.82	0.79			
	3. STEAM FEED lb/hr PSIG	NIL	NIL	NIL			
	4. GASIFIER PRESSURE PSIG	15	15	15			
	5.						
DESIGN DRIVERS	1. GASIFIER TEMP, °F						
	2.						
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW						
	2. H.P. STEAM, lb/lb	2.44	2.37	2.40			
	COAL FEED						
	3. M.P. STEAM, lb/HR						
	4. L.P. STEAM, lb/HR						
	5. BOILER FEEDWATER, GPM						
	6. COOLING WATER, GPM						

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 21
DESCRIPTION GAS COOLING

SHEET 1 OF 2

REFERENCE/CASE		2/MXL	3/	7/1	7/2	10/1 3/1	10/4 3/1
NO. OF PARALLEL TRAINS OPERATING/SPARE							
SYSTEM COST, MM \$		53.793 Mid 75	19.3 Oct 77	5.655 Mid 75	51.063* Mid 75	92.087 Mid 76	91.044 Mid 76
DATE OF MONEY							
COST DRIVERS	1. CAPACITY, Mlb/HR	4,172	2,178.5	1,375	1,364	1,708	1,708
	2. AGR TEMP, °F	NA	86	267	105		
	3. GAS TEMPERATURE, °F	1020	302	1100	160	394	394
	4. TYPE GASIFIER	BGC	Lurgi	Lurgi	K-T	Texaco	Texaco
	5.						
DESIGN DRIVERS	1. OIL & PARTICULATE REMOVAL		NA	NA	NA	NA	NA
	2. PRESSURE, PSIG		415	290	19.3		
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	NA	NA	102	NA	NA	NA
	2. H.P. STEAM, Mlb/HR (PSIG)		NA		-990.7 (1500)	NA	NA
	3. M.P. STEAM, Mlb/HR (PSIG)		NA			NA	NA
	4. L.P. STEAM, Mlb/HR	NA	NA			NA	NA
	5. BOILER FEEDWATER, GPM	NA	NA		2,080	NA	NA
	6. COOLING WATER, GPM	NA	NA	36,560	25,615	NA	NA

* Includes Cooler, Solids Handling, and Gasifier

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11	SYSTEM NUMBER
21	UNIT OPERATION NUMBER
GAS	DESCRIPTION

SHEET 2 OF 2

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 14

UNIT OPERATION NUMBER 22

DESCRIPTION ACID GAS REMOVAL - SELEXOL

SHEET 1 OF 4

A-B-12

REFERENCE/CASE	6/MX	6/PX	10/1	10/4	10/5	11/1	
NO. OF PARALLEL TRAINS	4	4	3	3		1	
OPERATING/SPARE							
SYSTEM COST, MM \$	30.773	25.152	22.263	62.758	19.441	5.440	
DATE OF MONEY	Mid 75	Mid 75	Mid 76	Mid 76	Mid 76	Apr 78	
COST DRIVERS	1. CAPACITY, FEEDFLOW, lb/HR	1,557,340	1,502,440	1,529,610	1,529,610	1,528,611	
	2. ABSORPTION PRESS, PSIG	242	255	335	335	1135	520
	3. H ₂ S/COS/CO ₂ REMOVED, MOL ON FEED	95/28/74	87/43/59	92/38/52	99.96/97.46/ 71.8	92/36/48.4	99/72/94
	4. SOLUTION LOSSES						
	5. REMOVAL OF H ₂ S/COS/CO ₂ (MOL/HR)	840/10.8/4522	794/15.2/1990	835/20.4/1164	907/52.9/3629	835/19.8/1018	17/1.09/437.38
DESIGN DRIVERS	1. CO ₂ CONTENT & REMOVAL REQ						
	2. MIN H ₂ S IN ACID GAS, %	17.64	27.26	37.3	17.9	39.7	3
	3. ABSORPTION TEMP, °F	103	103	105	57	105	91
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	NA	NA	NA	4873.2	NA	NA
	2. H.P. STEAM, lb/HR (PSIG						
	3. M.P. STEAM, lb/HR (PSIG				374,700 (100)		
	4. L.P. STEAM, lb/HR (PSIG	NA	209,000 (100)	102,300 (115)	326,100 (15)	5100 (115)	36,500 (145)
	5. BOILER FEEDWATER, GPM						
	6. COOLING WATER, GPM	11220	31200	7200	76,480	3120	

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 14

UNIT OPERATION NUMBER 22

DESCRIPTION ACID GAS REMOVAL - STRETFORD

SHEET 2 OF 4

COST DRIVERS	REFERENCE/CASE	4/					
	NO. OF PARALLEL TRAINS						
	OPERATING/SPARE						
	SYSTEM COST, MM \$	8.0					
DESIGN DRIVERS	DATE OF MONEY	4th Qtr 76					
	1. CAPACITY, TONS/DAY	3400					
	2. MOLES/HR OF SULFUR	59					
	3. HCN IN FEED GAS, %						
SYSTEM INTERFACES	4.						
	5.						
	1. CO ₂ IN FEED GAS, %						
	2. ABSORPTION PRESS, PSIG						
	3. ABSORPTION TEMP, °F						
	4. H ₂ S REMOVAL EFFICIENCY, %	99%					
	5. COS, %	70					
	6.						
	1. ELECTRICITY, KW						
	2. H.P. STEAM, lb/HR						
	3. M.P. STEAM, lb/HR						
	4. L.P. STEAM, lb/HR						
	5. BOILER FEEDWATER, GPM						
	6. COOLING WATER, GPM						

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 14
UNIT OPERATION NUMBER 22

DESCRIPTION ACID GAS REMOVAL - RECTISOL

SHEET 3 OF 4

REFERENCE/CASE	8/	9/2	17/				
NO. OF PARALLEL TRAINS		1/0					
OPERATING/SPARE							
SYSTEM COST, MM \$	48.633	31.060	115.136				
DATE OF MONEY	Mid 77	Mid 77	1st Qtr 78				
1. CAPACITY, SCFH	33,316,600	11,833,000	48,330,000				
2. PRESSURE, PSIG	370	440	355				
3. H ₂ S/COS REMOVAL							
4. SOLVENT LOSS							
5.							
COST DRIVERS							
1. H ₂ S/COS/CO ₂ FEED, %	1.0/0/40.9	1.1/NA/5.5	1.39/NA/32.5				
2. H ₂ S/COS/CO ₂ REMOVAL							
3. SULFUR RECOVERY METHOD							
4. ORGANICS/TRACE COMPONENTS							
5. IN FEED							
DESIGN DRIVERS							
1. ELECTRICITY, KW	10,614	5,820	15,660				
2. H.P. STEAM, lb/HR	0		248,100 (550)				
3. M.P. STEAM, lb/HR	0	5,292 (150)	14,100 (110)				
4. L.P. STEAM, lb/HR	97,763 (150)	66,150 (50)					
5. BOILER FEEDWATER, GPM	0	144	7,950				
6. COOLING WATER, GPM	12,810	12,000	9,000				
SYSTEM INTERFACES							

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 14
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - BENFIELD

SHEET 4 OF 4

REFERENCE/CASE	1*/1	1*/2	1*/3	1*/4	2/MSL	7/1	
NO. OF PARALLEL TRAINS	1/0	1/0	1/0	1/0	1/0	1/0	
OPERATING/SPARE							
SYSTEM COST, MM \$	15.584	14.181	12.779	18.389	NA	5.396	
DATE OF MONEY	Mid 76	Mid 76	Mid 76	Mid 76	Mid 75	Mid 75	
COST DRIVERS	1. CAPACITY, SCYH	5,206,250	7,631,670	6,319,170	10,402,080	34,949,000	17,490,000
	2. GAS PRESS, PSIG	395	395	395	395	265	275
	3. H ₂ S/COS REMOVAL REQ, PPM	≤50	≤50	≤50	≤50		
	4. SOLUTION LOSS						
	5.						
DESIGN DRIVERS	1. H ₂ S/COS/CO ₂ FEED, %	2.68/0.30/9.46	1.89/0.21/5.33	2.12/0.21/5.44	1.13/0.13/6.70	1.32/0.06/1.05	1.15/-/24.05
	2. SELECTIVITY						
	3. CO ₂ CONTENT, %						
	4. SULFUR RECOVERY METHOD						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	895	820	715	1245	NA	2,914
	2. H.P. STEAM, lb/HR	0	0	0	0	NA	
	3. M.P. STEAM, lb/HR	0	0	0	0	NA	
	4. L.P. STEAM, lb/HR (PSIG)	87,200 (60)	76,810 (60)	66,265 (60)	111,550 (60)	NA	
	5. BOILER FEEDWATER, GPM	G	0	0	0		
	6. COOLING WATER, GPM	110	100	90	150		78,390

* Benfield with Amine Scrubber

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 16
UNIT OPERATION NUMBER 23

SHEET 1 OF 1

DESCRIPTION		COMPRESSORS					
REFERENCE/CASE	1/1	€/EXL	€/EXL	6/EXL	8/		
NO. OF PARALLEL TRAINS		3/1	4/1				
OPERATING/SPARE							
SYSTEM COST, MM \$	77.29**	16.349	16.349	16.349	5.49		
DATE OF MONEY	Mid 76	Mid 75	Mid 75	Mid 75	Mid 77		
1. CAPACITY, MMSCFH*	5.5	9.2	15.6		0.59		
2. PRESSURE BOOST, PSI *	392	3.5	3.75		1040		
3. MATERIALS OF CONSTRUCTION							
4.							
5.							
COST DRIVERS							
DESIGN DRIVERS							
1. GAS PROPERTIES							
2. GASIFIER TYPE	K-T	Entrained Bed	Entrained Bed	Entrained Bed	K-T		
3.							
4.							
5.							
SYSTEM INTERFACES							
1. ELECTRICITY, KW	32,960						
2. H.P. STEAM, Mlb/HR (PSIG)	(-312.4) (600)	195.8 (1450)	220.7 (1450)		2100		
3. M.P. STEAM, Mlb/HR							
4. L.P. STEAM, Mlb/HR (PSIG)	9.2 (60)						
5. BOILER FEEDWATER, GPM							
6. COOLING WATER, GPM	6400						

* These cost drivers are also design drivers
** Cost and utilities include gasifiers

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 15
UNIT OPERATION NUMBER 31

DESCRIPTION SOLIDS TREATMENT SYSTEM

SHEET 1 OF 1

REFERENCE/CASE	3/1	16/Bituminous					
NO. OF PARALLEL TRAINS							
OPERATING/SPARE	5.6	28.312*					
SYSTEM COST, MM \$	Oct 77	1st Qtr 78					
DATE OF MONEY							
1. CAPACITY, Mlb/HR	244	353.27					
2.							
3.							
4.							
5.							
COST							
1. SLUDGE FEED CHARACTERISTICS	NA	NA					
2.							
3.							
4.							
5.							
DESIGN							
1. ELECTRICITY, KW	NA	NA					
2. H.P. STEAM, lb/HR	NA	NA					
3. M.P. STEAM, lb/HR	NA	NA					
4. L.P. STEAM, lb/HR	NA	NA					
5. BOILER FEEDWATER, GPM	NA	NA					
6. COOLING WATER, GPM	NA	NA					
SYSTEM INTERFACES							

* Cost includes Final Solids Disposal

SYSTEM NUMBER	UNIT OPERATION NUMBER	TAX
10	32	TAN

1001 127115

REFERENCE/CASE		3/1	7/1	17/				
NO. OF PARALLEL TRAINS								
OPERATING/SPARE								
SYSTEM COST, MR \$		18.6	7.6446	12.423				
DATE OF MONEY		Oct 77	Mid 75	1st Qtr 78				
1. CAPACITY, Mlb/NR*		1,662.7	263.1	2,079.3				
2. QUANTITY TAR/OIL,		36.391/51.711	32.6**	86.8/21.9				
3. Mlb/NR*								
4.								
5.								
COST DRIVERS								
DESIGN DRIVERS								
1. TREATMENT								
2.								
3.								
4.								
5.								
1. ELECTRICITY, KW		NA	429	1,659				
2. H.P. STEAM, Mlb/NR		NA						
3. H.P. STEAM, Mlb/NR		NA						
4. L.P. STEAM, Mlb/NR(PSIC)		NA		10.1(35)				
5. BOILER FEEDWATER, GPM		NA						
6. COOLING WATER, GPM		NA		3,860				
SYSTEM INTERFACES								

- These cost drivers are also design drivers

TAR PLUS OIL.

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 10
UNIT OPERATION NUMBER 33

PROCESS CONDENSATE TREATMENT										SHEET 1 OF 2	
DESCRIPTION	1/1	1/2	1/3	1/4	5/MACH	5/MISC	5/RANC				
REFERENCE/CASE											
NO. OF PARALLEL TRAINS	1/0	1/0	1/0	1/0	NA	NA	NA				
OPERATING/SPARE	2.304	2.304	2.165	2.534	59.067**	31.169**	9.83**				
SYSTEM COST, MM \$	Mid 76	Mid 76	Mid 76	Mid 76	Mid 76	Mid 76	Mid 76				
DATE OF MONEY	374	Prop	Prop	Prop	NA	NA	NA				
1. CAPACITY, GPM											
2. COMPOSITION OF STRIPPER											
3. EFFLUENT STRIPPING MEDIUM USED											
4. ANT OF STRIPPER REQ											
5.											
COST DRIVERS											
1. NO. OF EFFLUENT STREAMS											
2.											
3.											
4.											
5.											
DESIGN DRIVERS											
1. ELECTRICITY, KW	260	292	237	305	NA	NA	NA				
2. H.P. STEAM, lb/HR					NA	NA	NA				
3. M.P. STEAM, lb/HR					NA	NA	NA				
4. L.P. STEAM, lb/HR	39,758	40,568	36,193	46,309	NA	NA	NA				
5. BOILER FEEDWATER, GPM					NA	NA	NA				
6. COOLING WATER, GPM	145	150	130	170	NA	NA	NA				
SYSTEM INTERFACES											

* These cost drivers are also d. ign drivers

** Ref 5 cost includes Phenol Recovery

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 10

UNIT OPERATION NUMBER 33

DESCRIPTION PROCESS CONDENSATE TREATMENT

SHEET 2 OF 2

REFERENCE/CASE	6/MX	6/FX	11/1	16/Bituminous			
NO. OF PARALLEL TRAINS	NA	NA	NA	NA			
OPERATING/SPARE	NA	NA	NA	NA			
SYSTEM COST, MM \$	61.694	10.786	1.75	5.337			
DATE OF MONEY	Mid 75	Mid 75	Apr 78	1st Qtr 78			
COST DRIVERS							
1. CAPACITY, GPM	NA	NA	60	2644			
2. COMPOSITION OF STRIPPER EFFLUENT							
3. STRIPPING MEDIUM USED							
4. AMT OF STRIPPER REQ							
5.							
DESIGN DRIVERS							
1. NO. OF EFFLUENT STREAMS							
2.							
3.							
4.							
5.							
SYSTEM INTERFACES							
1. ELECTRICITY, KW	NA	NA	Prop	NA			
2. H.P. STEAM, lb/HR	NA	NA	Prop	NA			
3. M.P. STEAM, lb/HR	NA	NA	Prop	NA			
4. L.P. STEAM, lb/HR	NA	NA	Prop	NA			
5. BOILER FEEDWATER, GPM	NA	NA	Prop	NA			
6. COOLING WATER, GPM	NA	NA	Prop	NA			

* These cost drivers are also design drivers

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 10
UNIT OPERATION NUMBER 34
DESCRIPTION PHENOL RECOVERY

SHEET 1 OF 1

REFERENCE/CASE	1/1	1/2	1/3	1/4	7/1		
NO. OF PARALLEL TRAINS	1	1	1	1	NA		
OPERATING/SPARE							
SYSTEM COST, MM \$	3.00	3.00	2.82	3.33	8.387		
DATE OF MONEY	MID 76	MID 76	MID 76	MID 76	MID 75		
1. CAPACITY, lb/HR	PROP**	PROP**	PROP**	PROP**	3100 Phenol		
2. FEED STREAM CHARACTERISTICS							
3. P2O'D PHENOL/SOLVENT							
4. RECOVERY EFFICIENCY							
5. COST DRIVERS							
1. DESIGN DRIVERS							
2.							
3.							
4.							
5.							
SYSTEM INTERFACES							
1. ELECTRICITY, KW	168	172	153	193	328		
2. H.P. STEAM, lb/HR	0	0	0	0	0		
3. H.P. STEAM, lb/HR	0	0	0	0	0		
4. L.P. STEAM, lb/HR (PSIG)	-6,468	-6,600	-5,888	-7,534	76,600 (60)		
5. BOILER FEEDWATER, GPM	0	0	0	0	0		
6. COOLING WATER, GPM	550	560	500	610	1838		

* These cost drivers are also design drivers
** Proprietary unit; no mass balances given

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 10		UNIT OPERATION NUMBER 35		DESCRIPTION		MW, RECOVERY		3/1		16/Bituminous		17/																				SHEET 1 OF 1	
REFERENCE/CASE		NO. OF PARALLEL TRAINS		OPERATING/SPACE		SYSTEM COST, MM \$		DATE OF MONEY		10.0		11.006		1st Qtr 70		7.726		1st Qtr 70		1,077,571		0.48											
1. CAPACITY, lb/MR		2. AMT MW, IN P.E.D., %		3. STEAM REQ, lb/Mr, PSIG		4. CHEMICALS REQ		5.		1,559,101		0.63																					
DESIGN		1. OVERALL COMPOSITION OF		2. SYSTEM FEED		3.		4.		5.																							
SYSTEM INTERFACES		1. ELECTRICITY, KW		2. H.P. STEAM, Mlb/MR(PSIG)		3. H.P. STEAM, Mlb/MR(PSIG)		4. L.P. STEAM, Mlb/MR(PSIG)		5. BOILER FEEDWATER, GPM		6. COOLING WATER, GPM																					
		NA		NA		NA		540		15.1 (550)		78.1 (110)		172.9 (35)		5,160																	

* These cost drivers are also design drivers

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 18
UNIT OPERATION NUMBER 36
DESCRIPTION SULFUR RECOVERY

SHEET 1 OF 2

REFERENCE/CASE	1/1	1/2	1/3	1/4	7/2	7/1	
NO. OF PARALLEL TRAINS OPERATING/SPARE					3/0	3/0	
SYSTEM COST, MM \$	14.331	14.895	14.895	21.665	7.708	7.708	
DATE OF MONEY	Mid 76	Mid 76	Mid 76	Mid 76	Mid 75	Mid 75	
COST DRIVERS	1. CAPACITY, LT/DAY SULFUR	392	480	409	559	1298	
	2. H ₂ S IN FEED, %	37	44.9	41	36.4	20.39	
	3. REQ EFFICIENCY						
	4. FUEL REQ FOR INCINERATORS, MMBTU/HR						
	5. STEAM PRODUCED, lb/hr, PSIG				44,000/60	39,200/60	
DESIGN DRIVERS	1. AMT SOUR GAS AVAILABLE						
	2. SULFUR EMISSION REGULATIONS						
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	542	614	788	1166	386	
	2. H.P. STEAM, lb/HR	0	0	0	0	0	
	3. M.P. STEAM, lb/HR (PSIG)	0	0	0	0	0	
	4. L.P. STEAM, lb/HR (PSIG)	76,928(60)	93,900(60)	102,000(60)	153,700(60)	(-44,000)	(-39,200)
	5. BOILER FEEDWATER, GPM	162	197	214	323	117	88
	6. COOLING WATER, GPM	0	0	0	0	0	0

* These cost drivers are also design drivers

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 18
UNIT OPERATION NUMBER 36
DESCRIPTION SULFUR RECOVERY

SHEET 2 OF 2

A-B-24

REFERENCE/CASE		10/1	10/4					
COST DRIVERS	NO. OF PARALLEL TRAINS							
	OPERATING/SPARE	1/0	1/0					
	SYSTEM COST, MM \$	8.26	13.546					
	DATE OF MONEY	Mid 76	Mid 76					
COST DRIVERS	1. CAPACITY, LT/DAY SULPHUR	9.64	12.7					
	2. H ₂ S IN FEED, %	1.0	0.8					
	3. REQ EFFICIENCY							
	4. FUEL REQ FOR INCINERATORS, MMBTU/HR							
	5. STEAM PRODUCED, lb/hr, PSIG							
DESIGN DRIVERS	1. AMT SOUR GAS AVAILABLE							
	2. SULFUR EMISSION REGULATIONS							
	3.							
	4.							
	5.							
SYSTEM INTERFACES	1. ELECTRICITY, KW	NA	NA					
	2. H.P. STEAM, lb/HR	NA	NA					
	3. M.P. STEAM, lb/HR (PSIG)	NA	NA					
	4. L.P. STEAM, lb/HR (PSIG)	NA	NA					
	5. BOILER FEEDWATER, GPM	NA	NA					
	6. COOLING WATER, GPM	NA	NA					

* These cost drivers are also design drivers

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 10
UNIT OPERATION NUMBER 37
DESCRIPTION BIOLOGICAL TREATMENT

SHEET 1 OF 1

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REFERENCE/CASE	3/1	16/Bituminous					
NO. OF PARALLEL TRAINS OPERATING/SPARE							
SYSTEM COST, MM \$	25.4	5.657					
DATE OF MONEY	Oct 77	1st Qtr 78					
COST DRIVERS							
1. CAPACITY, Mlb/HR	1,983	1,340					
2. ZERO DISCHARGE REQ*							
3. CLARIFICATION REQ							
4. SURFACE MECHANICAL AERATORS							
DESIGN DRIVERS							
1. DEGREE OF REUSE							
2. POLLUTANT REMOVAL REQ							
3.							
4.							
5.							
SYSTEM INTERFACES							
1. ELECTRICITY, KW	NA	NA					
2. H.P. STEAM, lb/HR	NA	NA					
3. M.P. STEAM, lb/HR	NA	NA					
4. L.P. STEAM, lb/HR	NA	NA					
5. BOILER FEEDWATER, GPM	NA	NA					
6. COOLING WATER, GPM	NA	NA					

* These cost drivers are also design drivers

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 9
UNIT OPERATION NUMBER 39
DESCRIPTION COOLING WATER

SHEET 1 OF 1

REFERENCE/CASE	7/2	7/1	9/1	9/2	11/1		
NO. OF PARALLEL TRAINS	1/0	2/0					
OPERATING/SPARE							
SYSTEM COST, MM \$	1.703	3.392	1.888	6.371	5.06		
DATE OF MONEY	Mid 75	Mid 75	Mid 77	Mid 77	APR 78		
1. CAPACITY, gpm	38,000	92,000	25,940	173,190	55,100/15,000		
2. CHEMICAL TREATMENT RECOVERY							
3. SIDE STREAM TREATMENT							
4. CHEMICAL TREATMENT RECOVERY							
5.							
1. MAKEUP WATER QUALITY	Process	Process			Clarified		
2. CYCLES OF CONCENTRATION	NA	NA			NA		
3.							
4.							
5.							
1. ELECTRICITY, KW	4,431	8,442	1,116	6,060	NA		
2. H.P. STEAM, lb/HR					NA		
3. M.P. STEAM, lb/HR					NA		
4. L.P. STEAM, lb/HR					NA		
5. BOILER FEEDWATER, GPM					NA		
6. COOLING WATER, GPM					NA		

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 19
UNIT OPERATION NUMBER 41
DESCRIPTION INCINERATION

SHEET 1 OF 1

REFERENCE/CASE	17/						
	NO. OF PARALLEL TRAINS	3					
	OPERATING/SPARE						
	SYSTEM COST, MM \$	10.98					
COST DRIVERS	DATE OF MONEY	1st Qtr 78					
	1. CAPACITY, Mlb/HR	1,753					
	2. GAS PROPERTIES						
	3. INCINERATOR TYPE						
DESIGN DRIVERS	4. HEAT RECOVERY						
	5.						
	1. GAS TEMPERATURE, °F	1500					
	2. GAS FUEL VALUE						
SYSTEM INTERFACES	3.						
	4.						
	5.						
	1. ELECTRICITY, KW	980					
	2. H.P. STEAM, Mlb/HR						
	3. M.P. STEAM, Mlb/HR						
	4. L.P. STEAM, Mlb/HR	-99.5 (15)					
	5. BOILER FEEDWATER, GPM	205					
	6. COOLING WATER, GPM						

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 17		SHEET 2 OF 2									
UNIT OPERATION NUMBER 80											
DESCRIPTION AIR SEPARATION AND OXIDANT FEEDING											
REFERENCE/CASE	6/MX	7/2	7/1	9/2	11/	12/	17/				
NO. OF PARALLEL TRAINS	4/0										
OPERATING/SPARE	26.279	62.914	49.448	35.612	20.65	285.6	3/0				
SYSTEM COST, MM \$	Mid 75	Mid 75	Mid 75	Mid 77	Apr 78	4th Qtr 75	106.7				
DATE OF MONEY							1st Qtr 78				
1. CAPACITY, TONS/DAY	1,194	5,900	2,610	3,000	1,000	20,000	7,780				
2. PRODUCT PURITY %	98	98	98	98	99.5	98	98				
3. PRODUCT PRESSURE (PSIG)*	340	25	350		800	470	500				
4.											
5.											
COST DRIVERS											
1. PLANT SIZE											
2.											
3.											
4.											
5.											
DESIGN DRIVERS											
1. ELECTRICITY, KW	NA	3,291	37,000	26,300	120	400	(-2,004)				
2. H.P. STEAM, Mlb/HR(PSIG)	NA	1,042**			144.1(1150)	3,968(1200)	1,227(1500)				
3. M.P. STEAM, Mlb/HR(PSIG)	NA					(-2,604)(500)	6				
4. L.P. STEAM, Mlb/HR	NA					0	0				
5. BOILER FEEDWATER, GPM	NA					0	0				
6. COOLING WATER, GPM	NA	19,054,400	4,803,800	30,000	14,400	365,500	121,984				
SYSTEM INTERFACES											

* These cost drivers are also design drivers

** MMSTU/HR

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 17
UNIT OPERATION NUMBER 80

DESCRIPTION AIR SEPARATION AND OXIDANT FEEDING

SHEET 1 OF 2

REFERENCE/CASE	1/1	1/2	1/3	1/4	5/MISC	5/EXTC	5/EXTC-DP
NO. OF PARALLEL TRAINS	1/0	1/0	1/0	1/0	3/0	5/0	5/0
OPERATING/SPARE							
SYSTEM COST, MM \$	20.948	31.767	21.778	38.673	57.938	117.389	113.177
DATE OF MONEY	Mid 76	Mid 76	Mid 76	Mid 76	Mid 76	Mid 76	Mid 76
1. CAPACITY, TONS/DAY	1,720	2,160	1,785	3,030	5,100	8,379	7,876
2. PRODUCT PURITY %	99.5	99.5	99.5	99.5	98	98	98
3. PRODUCT PRESSURE (PSIG)	30	30	30	30	400	2	2
4.							
5.							
1. PLANT SIZE							
2.							
3.							
4.							
5.							
1. ELECTRICITY, KW	22,370	28,337	23,117	39,895	NA	NA	NA
2. H.P. STEAM, lb/HR	0	0	0	0	NA	NA	NA
3. M.P. STEAM, lb/HR	0	0	0	0	NA	NA	NA
4. L.P. STEAM, lb/HR	0	0	0	0	NA	NA	NA
5. BOILER FEEDWATER, GPM	0	0	0	0	NA	NA	NA
6. COOLING WATER, GPM	19,000	24,000	20,000	34,000	NA	NA	NA

* These cost drivers are also design drivers

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 3		UNIT OPERATION NUMBER 81		FINAL SOLIDS DISPOSAL		16/Bituminous 11/1		3.4		Oct 77		SHEET 1 OF 1	
DESCRIPTION		REFERENCE/CASE		NO. OF PARALLEL TRAINS		OPERATING/SPARE		SYSTEM COST, M \$		DATE OF MONEY			
1. CAPACITY, AREA FT ²		2. DISPOSAL METHOD		3. TRANSPORTATION		4.		5.					
COST		DESIGN		DRIVERS		1. CHARACTERISTICS OF		2. STORAGE RATE, Mlb/HR		52.565		244	
3.		4.		5.		1. ELECTRICITY, KW		2. H.P. STEAM, lb/HR		NA		NA	
SYSTEM		INTERFACES		3. M.P. STEAM, lb/HR		4. L.P. STEAM, lb/HR		5. BOILER FEEDWATER, GPM		NA		NA	
				6. COOLING WATER, GPM									

* Cost includes Solids Treatment System

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 6

UNIT OPERATION NUMBER 82

DESCRIPTION

BY-PRODUCT STORAGE AND LOADING

SHEET 1 OF 1

REFERENCE/CASE		7/1	3/1	17/				
NO. OF PARALLEL TRAINS OPERATING/SPARE								
SYSTEM COST, MM \$		1.685	22.5	5.26				
DATE OF MONEY		Mid 75	Oct 77	1st Qtr 78				
COST DRIVERS	1. CAPACITY, Mlb/HR	30.8/29/3.1	1,320.5/11.4					
	2. NO. OF BY-PRODUCTS*	TAR/NH ₃ /Phend	Mixed					
	3.							
	4.							
	5.							
DESIGN DRIVERS	1. QUANTITY STORED, MGBL			128.8				
	2.							
	3.							
	4.							
	5.							
SYSTEM INTERFACES	1. ELECTRICITY, KW	13	NA	60				
	2. H.P. STEAM, Mlb/HR		NA					
	3. H.P. STEAM, Mlb/HR(PSIG)		NA	9.6(110)				
	4. L.P. STEAM, Mlb/HR		NA					
	5. BOILER FEEDWATER, GPM		NA					
	6. COOLING WATER, GPM		NA					

* These cost drivers are also design drivers

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APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 18
UNIT OPERATION NUMBER 81
DESCRIPTION

SULFUR STORAGE AND LOADING

REFERENCE/CASE	1/1	1/2	1/3	1/4	9/1		
NO. OF PARALLEL TRAINS OPERATING/SPARE	3.306 Mid 76	3.637 Mid 76	3.373 Mid 76	4.099 Mid 76	0.174 Mid 77		
SYSTEM COST, MM \$	725.8	887.0	806.4	1,048.3	NA		
DATE OF MONEY	435.6	513.3	454.3	620.9	30 58.0		
1. LOADING CAPACITY, MID/Hr							
2. STORAGE TIME REQ'D, DAYS							
3. SULFUR PRODUCTION, LT/DAY							
4. COOLERS							
5. DRIVERS							
1. AMBIENT TEMP RANGE, °F							
2. DRIVERS							
3. DRIVERS							
4. DRIVERS							
5. DRIVERS							
1. ELECTRICITY, KW	154	188	173	223	5		
2. H.P. STEAM, lb/hr	0	0	0	0	0		
3. M.P. STEAM, lb/hr	0	0	0	0	0		
4. L.P. STEAM, lb/hr	0	0	0	0	0		
5. BOILER FEEDWATER, GPM	0	0	0	0	0		
6. COOLING WATER, GPM	0	0	0	0	0		

SHEET 1 OF 1

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 8

UNIT OPERATION NUMBER 84

DESCRIPTION STEAM GENERATION

SHEET 1 OF 2

REFERENCE/CASE	5/MACW	5/MXSC	6/MX	7/2	10/1	10/4	10/5
NO. OF PARALLEL TRAINS OPERATING/SPARE	6/0	9/0	3	1			
SYSTEM COST, MM \$	41.987	2.057	68.485	4.0	1.092	1.092	1.092
DATE OF MONEY	Mid 76	Mid 76		Mid 75	Mid 76	Mid 76	Mid 76
COST DRIVERS	1. CAPACITY, MIB/HR	2261.2	2830.6	1711.0	990.7	3515.5	3460.7
	2. STEAM PRESSURE, PSIG *	1450	1450	1450	NA	1450	1450
	3. DEGREE OF SUPERHEAT*						
	4. FLUE GAS TREATMENT REQ'D*						
	5. BOILER TYPE						
DESIGN DRIVERS	1. STEAM REQUIREMENTS						
	2. COAL PROPERTIES						
	3.						
	4.						
	5.						
SYSTEM DESIGN PARAMETERS	1. ELECTRICITY, KW				-20,000		
	2. H.P. STEAM, MIB/HR	-2261.2	-2830.6	-1711.0	990.7	-3515.5	-3460.7
	3. M.P. STEAM, MIB/HR	-258.1	-179.6	-395.1		-35.0	-69.7
	4. L.P. STEAM, MIB/HR					-87.5	-359.6
	5. BOILER FEEDWATER, GPM	8100	7213	3700	2080	9296	9296
	6. COOLING WATER, GPM						9765

* These cost drivers are also design drivers

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 8

UNIT OPERATION NUMBER 84

DESCRIPTION STEAM GENERATION

SHEET 2 OF 2

REFERENCE/CASE		11/1						
NO. OF PARALLEL TRAINS								
OPERATING/SPARE								
SYSTEM COST, MM \$		1.190						
DATE OF MONEY		4/78						
COST DRIVERS	1. CAPACITY, MLD/HR	167.6						
	2. STEAM PRESSURE (PSIG)*	1175						
	3. DEGREE OF SUPERHEAT*							
	4. FLUE GAS TREATMENT REQ'D*							
	5. BOILER TYPE							
DESIGN DRIVERS	1. STEAM REQUIREMENTS							
	2. COAL PROPERTIES							
	3.							
	4.							
	5.							
SYSTEM INTERFACES	1. ELECTRICITY, KW							
	2. H.P. STEAM, MLD/HR	-467.6						
	3. H.P. STEAM, MLD/HR	-57.12						
	4. L.P. STEAM, MLD/HR	-49.6						
	5. BOILER FEEDWATER, GPM	1695						
	6. COOLING WATER, GPM							

* These cost drivers are also design drivers

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**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 7
UNIT OPERATION NUMBER 85
DESCRIPTION RAW WATER TREATMENT

SHEET 1 OF 2

REFERENCE/CASE	1/1	1/2	1/3	1/4	9/1	9/2	11/
NO. OF PARALLEL TRAINS OPERATING/SPARE							
SYSTEM COST, MM \$	17.801*	21.718*	18.87*	25.992*	6.899	20.562	4.960
DATE OF MONEY	Mid 76	Mid 76	Mid 76	Mid 76	Mid 77	Mid 77	Apr 78
COST DRIVERS 1. CAPACITY, GPM 2. 3. 4. 5.					1,170	6,726	1,393
DESIGN DRIVERS 1. STEAM PRESSURE, PSIG 2. RAW WATER QUALITY 3. 4. 5.					RIVER	RIVER	WELL WATER TDS: 300ppb Silica: 20ppb
SYSTEM INTERFACES 1. ELECTRICITY, KW 2. H.P. STEAM, MB/HR (PSIG) 3. H.P. STEAM, LB/HR 4. L.P. STEAM, MB/HR (PSIG) 5. BOILER FEEDWATER, GPM 6. COOLING WATER, GPM	23,000 (-159.4) (600)	30,600 (-155.1) (600)	25,900 (-415.8) (600)	39,600 (-381.3) (600)	62	350	288.4

* Cost and Utilities Include Power Generation

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APPENDIX B

2 OF 2

DESCRIPTION	RAM WATER TREATMENT
FIELDPIECE/CASE NO. OF PARALLEL TRAINS OPERATING/SPARE SYSTEM COST, RM \$ DATE OF WORKY	16, BITUMENS 16,096 1st Oct 78 14, 140
1. CAPACITY, GPM 2. 3. 4. 5.	
1. STEAM PRESSURE, PSIG 2. RAM WATER QUALITY 3. 4. 5.	
1. ELECTRICITY, KW 2. H.P. STEAM, MH/HOUR-SEC 3. H.P. STEAM, MH/HIP 4. L.P. STGZ. MH/HOUR-SEC 5. BOILER EFFICIENCY, GPH 6. COOLING WATER, GPH	NA

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 8
UNIT OPERATION NUMBER 86
DESCRIPTION FLOUE GAS TREATMENT - WELLMAN LORD

SHEET 1 OF 2

A-B-37

REFERENCE/CASE	16/Bituminous						
NO. OF PARALLEL TRAINS							
OPERATING/SPARE							
SYSTEM COST, MM \$	96.001						
DATE OF MONEY	1st Qtr 78						
DESIGN DRIVERS	1. CAPACITY, * SCFH 2. SO ₂ REMOVAL REQ, % 3. AMT SO ₂ /SO ₃ REMOVAL 4. GAS REHEAT AT REQ'D, °F 5.	61,420,000 95					
DESIGN DRIVERS	1. SO ₂ /SO ₃ IN FLUE GAS, 2. HCl/ASH/OTHER, PPM/PPM/PPM 3. O ₂ , % 4. 5.	314/0 0/19/0					
SYSTEM INTERFACES	1. ELECTRICITY, KW 2. H.P. STEAM, lb/HR(PSIG) 3. M.P. STEAM, lb/HR 4. L.P. STEAM, lb/HR(PSIG) 5. BOILER FEEDWATER, GPM 6. COOLING WATER, GPM						

* Flue gas & vent gas feed to unit

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 8
UNIT OPERATION NUMBER 86
DESCRIPTION FLUE GAS TREATMENT - DOUBLE ALKALI

SHEET 2 OF 2

REFERENCE/CASE NO. OF PARALLEL TRAINS OPERATING/SPARE SYSTEM COST, MM \$ DATE OF MONEY	21						
	2/0						
	17.2						
	1977						
COST DRIVERS	1. CAPACITY, * SCFH	21.0 x 10 ⁶					
	2. SO ₂ REMOVAL REQ, %	94.2					
	3. AMT SO ₂ /SO ₃ REMOVAL	184 lb/MIN					
	4. GAS REHEAT AT REQ'D, °F	50					
	5.						
DESIGN SPECIFICATIONS	1. SO ₂ /SO ₃ IN FLUE GAS, PPM/PPM	34/1/1					
	2. HCl/ASH/OTHER, PPM/PPM/PPM	-					
	3. O ₂ , %	5.7					
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	1551					
	2. H.P. STEAM, lb/HR (PSIG)	NA					
	3. M.P. STEAM, lb/HR	NA					
	4. L.P. STEAM, lb/HR	NA					
	5. BOILER FEEDWATER, GPM						
	6. COOLING WATER, GPM	NA					

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SHEET 1 OF 1

REFERENCE/CASE	NO. OF PARALLEL TRAINS OPERATING/SPARE	SYSTEM COST, RM \$	DATE OF BIDDY	3/1	11/1	16/BID DATES
1. CAPACITY, KM	10.3	Oct 77	51,400	6.78	32,000	12.04B
2. VOLTAGE LEVEL, KV *			6	Apr 78	230/15/4.16	1st Qtr 78
3. PEAK CURRENT DEMAND, AMPS					1200	NA
4. NO. OF SERVICES						
5.						
1. RELIABILITY FACTOR						
2. MOTOR HORSEPOWER						
3.						
4.						
5.						
1. ELECTRICITY, KW				0		NA
2. H.P. STEAM, MIL/HR				0		NA
3. M.P. STEAM, MIL/HR				0		NA
4. L.P. STEAM, MIL/HR				0		NA
5. BOILER FEEDWATER, GPH				0		NA
6. COOLING WATER, GPM				0		NA

• **Those cost drivers are also design drivers**

APPENDIX B
SYSTEM CHARACTERIZATION

SYSTEM NUMBER 1
UNIT OPERATION NUMBER 88
DESCRIPTION GENERAL FACILITIES

SHEET 1 OF 1

REFERENCE/CASE NO. OF PARALLEL TRAINS OPERATING/SPARE SYSTEM COST, MM \$ DATE OF MONLY	1/1	1/2	1/3	1/4	9/1	9/2	
1. NO. OF BUILDINGS	61.1 Mid 76	65.5 Mid 76	62.8 Mid 76	70.8 Mid 76	2.06 Mid 77	3.34 Mid 77	
2. INCLUDED SERVICES							
3. TRANSPORTATION REQ.							
4.							
5.							
1. PERSONNEL PROTECTION							
2. PERSONNEL EGRESS							
3.							
4.							
5.							
1. ELECTRICITY, KW	NA	NA	NA	NA	440	231	
2. H.P. STEAM, lb/HR	NA	NA	NA	NA			
3. M.P. STEAM, lb/HR	NA	NA	NA	NA			
4. L.P. STEAM, lb/HR	NA	NA	NA	NA			
5. BOILER FEEDWATER, GPM	NA	NA	NA	NA			
6. COOLING WATER, GPM	NA	NA	NA	NA			

**APPENDIX B
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER 4		CONTROL AND INSTRUMENTATION										SHEET 1 OF 1
UNIT OPERATION NUMBER 89												
DESCRIPTION												
REFERENCE/CASE												
NO. OF PARALLEL TRAINS												
OPERATING/SPARE												
SYSTEM COST, MM \$												
DATE OF MONEY												
1.												
2.												
3.												
4.												
5.												
COST DRIVERS												
1.												
2.												
3.												
4.												
5.												
DESIGN DRIVERS												
1.												
2.												
3.												
4.												
5.												
SYSTEM INTERFACES												
1. ELECTRICITY, KW												
2. H.P. STEAM, lb/hr												
3. M.P. STEAM, lb/hr												
4. L.P. STEAM, lb/hr												
5. BOILER FEEDWATER, GPM												
6. COOLING WATER, GPM												

APPENDIX C

SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

UNIT OPERATION NUMBER		PAGE
10	COAL RECEIVING, STORAGE AND TRANSFER	C-1
11	COAL PREPARATION AND FEEDING	C-2
20	GASIFICATION	C-3
21	GAS COOLING	C-8
22	ACID GAS REMOVAL	C-9
23	COMPRESSION	C-13
31	SOLIDS TREATMENT SYSTEM	C-14
32	TAR-OIL SEPARATION	C-15
33	PROCESS CONDENSATE TREATMENT	C-16
34	PHENOL RECOVERY	C-17
35	AMMONIA RECOVERY	C-18
36	SULFUR RECOVERY	C-19
37	BIOLOGICAL TREATMENT	C-20
39	COOLING WATER SYSTEM	C-21
41	INCINERATION	C-22
80	AIR SEPARATION AND OXIDANT FEEDING	C-23
81	FINAL SOLIDS DISPOSAL	C-24
82	BY-PRODUCT STORAGE AND LOADING	C-25
83	SULFUR STORAGE AND LOADING	C-26
84	STEAM GENERATION	C-27
85	RAW WATER TREATMENT	C-28
86	FLUE GAS TREATMENT	C-29
87	PLANT ELECTRICAL SYSTEM	C-30
88	BUILDINGS AND SUPPORT FACILITIES	C-31
89	CONTROL AND INSTRUMENTATION	C-32

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 2

UNIT OPERATION NUMBER 10

DESCRIPTION COAL RECEIVING, STORAGE, & TRANSFER

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	COAL			
	ORIGIN	BATTERY LIMIT			
OUTPUTS	MAJOR COMPONENTS	C, H, S, O, N			
	SIG. MINOR COMPONENTS	ASH, H ₂ O			
	TEMPERATURE, °F	AMBIENT			
	PRESSURE, PSIG	ATMOSPHERIC			
	DESCRIPTION	COAL	REFUSE	RUN-OFF	
	DESTINATION	UNIT OP. 11	UNIT OP. 81	UNIT OP. 85	
BY-PRODUCTS	MAJOR COMPONENT	C, H, S, O, N	ROCK	H ₂ O	
	SIG. MINOR COMPONENT	ASH, WATER	---	COAL DUST	
	TEMPERATURE, °F	AMBIENT	AMBIENT	AMBIENT	
	PRESSURE, PSIG	ATMOSPHERIC	ATMOSPHERIC	ATMOSPHERIC	
	DESCRIPTION				
	USES				
RAW MATERIALS	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
	1.				
	2.				
	3.				
	4.				
5.					
6.					
7.					
8.					

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 11

UNIT OPERATION NUMBER 11

DESCRIPTION COAL PREPARATION AND FEEDING

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		COAL UNIT OP. 10 C, H, S, O, N ASH, H ₂ O AMBIENT ATMOSPHERIC			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SIZED COAL UNIT OP. 20 C, H, S, O, N ASH, H ₂ O AMBIENT ATMOSPHERIC	COAL FINES UNIT OP. 84 COAL DUST ASH, H ₂ O AMBIENT ATMOSPHERIC	DRYER OFF GAS ATMOSPHERIC AIR CO ₂ , H ₂ O 180°F ATMOSPHERIC	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		COAL FINES UNIT OP. 82 COAL DUST ASH, H ₂ O AMBIENT ATMOSPHERIC			
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION-KOPPERS-TOTZEK

SHEET 1 OF 5

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SIZED COAL UNIT OP. 11 C, H, S, O, N ASH, H ₂ O AMBIENT ATMOSPHERIC	OXYGEN UNIT OP. 80 O ₂ N ₂ , Ar AMBIENT 15	STEAM UNIT OP. 84 H ₂ O --- 250-900 15-1500	BFW UNIT OP. 85 H ₂ O --- 250 15-1500
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 21 H ₂ , CO ₂ , CO, H ₂ O H ₂ S, CO ₅ , N ₂ 600 15	ASH & SLAG UNIT OP. 31 SLAG H ₂ O 100-300 ATMOSPHERIC	STEAM UNIT OP. 84 H ₂ O --- 900 1450	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SLAG LANDFILL SLAG H ₂ O, ASH AMBIENT ATMOSPHERIC			
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 12

UNIT OPERATION NUMBER 20

DESCRIPTION GASIFICATION - TEXACO

SHEET 2 OF 5

		NO. 1	NO. 2		NO. 3
INPUTS	DESCRIPTION	SIZED COAL	OXYGEN		BFW
	ORIGIN	UNIT OP. 11	UNIT OP. 80		UNIT OP. 85
	MAJOR COMPONENTS	C, H, S, O, N	O ₂		H ₂ O
	SIG. MINOR COMPONENTS	ASH, H ₂ O	N ₂ , Ar		---
	TEMPERATURE, °F	100-200	300		250
	PRESSURE, PSIG	775	775		
OUTPUTS	DESCRIPTION	NO. 1	NO. 2		NO.
	DESTINATION	RAW GAS	ASH		
	MAJOR COMPONENT	UNIT OP. 20	UNIT OP. 31		
	SIG. MINOR COMPONENT	H ₂ , CO, CO ₂ , H ₂ O	ASH, WATER		
	TEMPERATURE, °F	H ₂ S, COS, N ₂	SOOT		
	PRESSURE, PSIG	300-400	150+		
		635	ATMOSPHERIC		
BY-PRODUCTS	DESCRIPTION	NO. 1	NO. 2		NO.
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - LURGI

SHEET 3 OF 5

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SIZED COAL UNIT OP. 11 C, H, S, N, O ASH, H ₂ O AMBIENT ATMOSPHERIC	OXYGEN UNIT OP. 80 O ₂ N ₂ , Ar 200 360	STEAM UNIT OP. 84 H ₂ O --- 650 450	RECYCLE LIQUOR UNIT OP. 21 PHENOL, TAR H ₂ O, NH ₃ 200 400
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 21 H ₂ , CO, CH ₄ , H ₂ S OILS, TAR, PHENOL 1100-800 300	SLAG/ASH UNIT OP. 31 H ₂ O, SLAG ASH AMBIENT ATMOSPHERIC	LOCK GAS UNIT OP. 21 CO ₂ , H ₂ , H ₂ O, CO H ₂ S, ASH 200 ATMOSPHERIC	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1.	ORIGINAL PAGE IS OF POOR QUALITY			
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION-BRITISH GAS CORP-LURGI

SHEET 4 OF 5

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SIZED COAL UNIT OP. 11 C, H, O, S, N ASH, H ₂ O AMBIENT ATMOSPHERIC	OXYGEN UNIT OP. 80 O ₂ N ₂ , Ar 200 360	STEAM UNIT OP. 84 H ₂ O --- 650 450	RECYCLE LIQUO UNIT OP. 21 PHENOL, TAR OIL, NH ₃ , H ₂ O 200 400
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 21 H ₂ , CO, CH ₄ , H ₂ O H ₂ S, TAR, OIL 800-350 300	LOCK GAS UNIT OP. 21 H ₂ , CO, CO ₂ H ₂ O 200 ATMOSPHERIC	SLAG/ASH UNIT OP. 31 SLAG ASH, H ₂ O AMBIENT ATMOSPHERIC	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SLAG CONSTRUCTION SLAG ASH, H ₂ O AMBIENT ATMOSPHERIC			
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 12
UNIT OPERATION NUMBER 20
DESCRIPTION GASIFICATION - BABCOCK & WILCOX

SHEET 5 OF 5

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		COAL UNIT OP. 11 C, H, O, S, N ASH, H ₂ O AMBIENT ATMOSPHERIC	OXYGEN UNIT OP. 80 O ₂ N ₂ , Ar AMBIENT 275	BFW UNIT OP. 84 H ₂ O --- 250 400-1500	STEAM UNIT OP. 85 H ₂ O --- 400 225
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 21 CO ₂ , H ₂ CO ₂ , H ₂ O 1800 225	STEAM UNIT OP. 84 H ₂ O --- 900 1500	SLAG UNIT OP. 31 SLAG ASH, H ₂ O 100 ATMOSPHERIC	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SLAG CONSTRUCTION SLAG H ₂ O, ASH AMBIENT ATMOSPHERIC			
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 13
UNIT OPERATION NUMBER 21
DESCRIPTION GAS COOLING - (NOTES)

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 20 H ₂ , CO, CO ₂ , H ₂ O H ₂ S, COS, NH ₃ 1800-400 15-650	MAKE UP WATER UNIT OP. 85 H ₂ O --- AMBIENT 15-650		
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 22 H ₂ , CO, CO ₂ , H ₂ O H ₂ S, COS, NH ₃ 100-200 5-600	PROCESS COND UNIT OP. 31 H ₂ O ASH 100-200 100	GAS LIQUOR UNIT OP. 33 H ₂ O, ASH, OIL PHENOL, NH ₃ 280 270	STEAM UNIT OP. 84 H ₂ O - 598 1500
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. SULFURIC ACID 2. SODIUM HYDROXIDE 3. 4. 5. 6. 7. 8.	NOTES:			
		STREAM Nos. 1&2 APPLY TO K-T, TEXACO AND B&W. STREAM Nos. 1&3 APPLY TO LURGI AND BGC LURGI.			

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 14
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - SELEXOL

SHEET 1 OF 4

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 21 H ₂ , CO, CO ₂ , H ₂ O H ₂ S, CO ₂ , HCN 100-150 300-700			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SWEET GAS UNIT OP. 23 H ₂ , CO, CO ₂ CH ₄ 100-150 300-700	ACID GAS UNIT OP. 36 H ₂ S, COS CO ₂ , H ₂ O 150-200 10	PURGE H ₂ O UNIT OP. 33 H ₂ O SOLVENT 150-200 ATMOSPHERIC	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. DIMETHYL ETHER OF POLYETHYLENE GLYCOL 2. 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 14
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - STRETFORD

SHEET 2 OF 4

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RAW GAS UNIT OP. 21 H ₂ , CO, CO ₂ , H ₂ O H ₂ S, COS, CH ₄ 120 15-17	AIR BATTERY LIMIT O ₂ , N ₂ --- AMBIENT ATMOSPHERIC		
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SWEET GAS UNIT OP. 23 H ₂ , CO, CO ₂ , H ₂ O CH ₄ 120 1-2	VENT GAS ATMOSPHERE O ₂ , N ₂ NH ₃ 100-150 ATMOSPHERIC	PURGE LIQUOR UNIT OP. 33 SOLVENT THIOCYANATE 150-300 10	
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SULFUR SALES S --- 280 ATMOSPHERIC			
RAW MATERIALS	1. ANTHRAQUINONE DISULFURIC ACID 2. VANADIUM PENTOXIDE 3. ETHYLENE DIAMINE TETRA ACETIC ACID 4. SODIUM CARBONATE 5. SODIUM BICARBONATE 6. IRON SALTS 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 14

UNIT OPERATION NUMBER 22

DESCRIPTION ACID GAS REMOVAL - RECTISOL

SHEET 3 OF 4

		NO. 1	NO. 2	NO. 3	NO. 4
	INPUTS	CRUDE GAS UNIT OP. 21 H ₂ , CO, CO ₂ , H ₂ O H ₂ S, COS, NH ₃ 80-100 300			
	OUTPUTS	NO. 1	NO. 2	NO. 3	NO. 4
		SWEET GAS UNIT OP. 23 H ₂ , CO, CO ₂ CH ₄ , H ₂ O	ACID GAS UNIT OP. 36 H ₂ S, COS, CO ₂ , H ₂ O	PROCESS COND UNIT OP. 33 H ₂ O H ₂ S, CO ₂	
	BY-PRODUCTS	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. METHANOL 2. REFRIGERANT (FREON, AMMONIA, PROPYLENE) 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 14
UNIT OPERATION NUMBER 22
DESCRIPTION ACID GAS REMOVAL - BENFIELD

SHEET 4 OF 4

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		CRUDE GAS UNIT OP. 21 H ₂ , CO, CO ₂ , H ₂ O H ₂ S, OIL, NH ₃ 200 270-650			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SWEET GAS UNIT OP. 23 H ₂ , CO, CO ₂ , H ₂ O CH ₄ , H ₂ S	ACID GAS UNIT OP. 36 H ₂ S, COS CO ₂ , H ₂ O		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. POTASSIUM CARBONATE 2. ACTIVATOR 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 16

UNIT OPERATION NUMBER 23

DESCRIPTION COMPRESSION

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	SWEET GAS			
	ORIGIN	UNIT OP. 22			
	MAJOR COMPONENTS	H ₂ , CO, CO ₂			
	SIG. MINOR COMPONENTS	H ₂ O, H ₂ S			
	TEMPERATURE, °F	100-200			
	PRESSURE, PSIG	15-300			
OUTPUTS	DESCRIPTION	SALES GAS			
	DESTINATION	BATTERY LIMIT			
	MAJOR COMPONENT	H ₂ , CO, CO ₂			
	SIG. MINOR COMPONENT	H ₂ O, H ₂ S			
	TEMPERATURE, °F	100-150			
	PRESSURE, PSIG	600			
BY-PRODUCTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 15
UNIT OPERATION NUMBER 31
DESCRIPTION SOLIDS TREATMENT SYSTEM

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SLAG/ASH UNIT OP. 20 SLAG, H ₂ O ASH AMBIENT ATMOSPHERIC	BIO. SLUDGE UNIT OP. 37 BIO. SOLIDS H ₂ O AMBIENT ATMOSPHERIC	LIME SLUDGE UNIT OP. 85 CaCO ₃ , MgCO ₃ H ₂ O AMBIENT ATMOSPHERIC	BLOWDOWNS UNIT OP. 84,3 Ca, Mg, H ₂ O Zn, Cr, Cl, SiO ₂ AMBIENT ATMOSPHERIC
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		DEWATERED SLDS UNIT OP. 81 SOLIDS H ₂ O AMBIENT ATMOSPHERIC	TREAT. WATER UNIT OP. 33 H ₂ O TRACE SALTS AMBIENT ATMOSPHERIC		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. DEWATERING AIDS 2. THICKENING CHEMICALS 3. ANTIFOAM & SCALING CHEMICALS 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 10
UNIT OPERATION NUMBER 32
DESCRIPTION TAR-OIL SEPARATION

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	QUENCH H ₂ O	SOUR WATER		
	ORIGIN	UNIT OP. 21	UNIT OP. 22		
	MAJOR COMPONENTS	TAR, OIL	PHENOL, H ₂ O		
	SIG. MINOR COMPONENTS	H ₂ O, DUST	OILS		
	TEMPERATURE, °F	100-200	100-200		
	PRESSURE, PSIG	0-300	200		
OUTPUTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	DESTINATION				
	MAJOR COMPONENT	TAR, OIL	PHENOL, H ₂ O		
	SIG. MINOR COMPONENT	UNIT OP. 82,33	UNIT OP. 34		
	TEMPERATURE, °F	OILS, TAR	PHENOL		
	PRESSURE, PSIG	H ₂ O, H ₂ S	H ₂ O		
BY-PRODUCTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	USES				
	MAJOR COMPONENT	TAR, OIL			
	SIG. MINOR COMPONENT	COMBUSTION			
	TEMPERATURE, °F	TAR, OIL			
	PRESSURE, PSIG	H ₂ O			
RAW MATERIALS	1.	AMBIENT			
	2.	ATMOSPHERIC			
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 10

UNIT OPERATION NUMBER 33

DESCRIPTION PROCESS CONDENSATE TREATMENT

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	SOUR CONDENSATE	SOLID TRT H ₂ O		
	ORIGIN	UNIT OP. 20,21	UNIT OP. 31		
	MAJOR COMPONENTS	H ₂ O, H ₂ S	H ₂ O, H ₂ S		
	SIG. MINOR COMPONENTS	NH ₃ , OIL	SOLIDS		
	TEMPERATURE, °F	100-200	AMBIENT		
	PRESSURE, PSIG	0-100	ATMOSPHERIC		
OUTPUTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	DESTINATION	TREATED H ₂ O	ACID GAS		
	MAJOR COMPONENT	UNIT OP. 35,39	UNIT OP. 22		
	SIG. MINOR COMPONENT	H ₂ O	H ₂ S, NH ₃		
	TEMPERATURE, °F	ORGANICS			
	PRESSURE, PSIG	220	220		
BY-PRODUCTS	DESCRIPTION	15-100	15-100		
	USES	NO. 1	NO. 2	NO. 3	NO. 4
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	1. SULFURIC ACID				
	2. SODIUM HYDROXIDE				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 10

UNIT OPERATION NUMBER 34

DESCRIPTION PHENOL RECOVERY

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		PHENOLIC COND. UNIT OP. 21 PHENOL, H ₂ O OIL, TAR --- ---	STEAM UNIT OP. 84 H ₂ O --- 400 250		
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		TREATED H ₂ O UNIT OP. 21,35 H ₂ O PHENOL, NH ₃ 100-200 ---	CRUDE PHENOL UNIT OP. 84 PHENOL H ₂ O 100-200 100		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		CRUDE PHENOL SALES PHENOL H ₂ O AMBIENT ATMOSPHERIC			
RAW MATERIALS	1. SOLVENT 2. 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 10
UNIT OPERATION NUMBER 35
DESCRIPTION AMMONIA RECOVERY

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		NH ₃ FEED UNIT OP. 33,34 NH ₃ , CO ₂ H ₂ S, PHENOL --- ---	STEAM UNIT OP. 84 H ₂ O - 600 550		
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		STRIPPER BTMS UNIT OP. 39,37 H ₂ O NH ₃ , H ₂ S 100-200 ---	SOUR GAS UNIT OP. 36 H ₂ S NH ₃ 100-200 ---		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		AMMONIA SALES NH ₃ H ₂ O AMBIENT ---			
RAW MATERIALS	1. PHOSPHORIC ACID 2. SODIUM HYDROXIDE 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 18
UNIT OPERATION NUMBER 36
DESCRIPTION SULFUR RECOVERY

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SOUR GAS UNIT OP 33,34,35 H ₂ S, COS CO ₂ , NH ₃ 100-200 1-10	TAIL GAS UNIT OP. 36 SO ₂ H ₂ S 200-300 1-5		
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		TAIL GAS ATMOSPHERE N ₂ , CO ₂ , H ₂ O SO ₂ 280 ATMOSPHERIC	SULFUR UNIT OP. 83 SULFUR --- 240 ATMOSPHERIC		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SULFUR SALES SULFUR --- 240 ATMOSPHERIC			
RAW MATERIALS	1. CLAUS UNIT CATALYST 2. BEAVON UNIT CATALYST 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 10

UNIT OPERATION NUMBER 37

DESCRIPTION BIOLOGICAL TREATMENT

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SANITARY WASTE BATTERY LIMIT SEWAGE H ₂ O AMBIENT ATMOSPHERIC	SOUR WATER UNIT OP. 33 H ₂ O H ₂ S, OIL AMBIENT ATMOSPHERIC	NH ₃ WATER UNIT OP. 35 H ₂ O H ₂ S, NH ₃ AMBIENT ATMOSPHERIC	OILY WATER UNIT OP. 21,3 H ₂ O, ASH, H ₂ S PHENOL, OIL AMBIENT ATMOSPHERIC
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		BIO SLUDGE UNIT OP. 31 SOLIDS H ₂ O AMBIENT ATMOSPHERIC	TREATED H ₂ O UNIT OP. 39, 85 H ₂ O TRACE SALTS AMBIENT 0-100		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. SULFURIC ACID 2. SODIUM HYDROXIDE 3. PHOSPHORIC ACID 4. POLYMERS 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 9
UNIT OPERATION NUMBER 39
DESCRIPTION COOLING WATER SYSTEM

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
	INPUTS	HOT WATER BATTERY LIMIT H ₂ O DISS. SOLIDS 90-120 30-60			
	OUTPUTS	NO. 1	NO. 2	NO. 3	NO. 4
		COLD WATER BATTERY LIMIT H ₂ O ADDITIVES 70-90 60-100			
	BY-PRODUCTS	NO. 1	NO. 2	NO. 3	NO. 4
		CHROMATE REUSE CHROMATE ZINC 90-120 60-100			
RAW MATERIALS	1. SULFURIC ACID 2. CHLORINE 3. ALGECIDES 4. POLYPHOSPHATES 5. ZINC INHIBITOR 6. CHROME INHIBITOR 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 19
UNIT OPERATION NUMBER 41
DESCRIPTION INCINERATION

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	LOCK GAS	TAIL GAS	VENT GAS	
	ORIGIN	UNIT OP. 20	UNIT OP. 36	UNIT OP. 82	
	MAJOR COMPONENTS	H ₂ S, CO	H ₂ S, SO ₂	HYDROCARBON	
	SIG. MINOR COMPONENTS	HYDROCARBON	---	---	
	TEMPERATURE, °F	AMBIENT	AMBIENT	AMBIENT	
	PRESSURE, PSIG	1-5	1-5	0.5-1.0	
OUTPUTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	DESTINATION	FLUE GAS			
	MAJOR COMPONENT	ATMOSPHERE			
	SIG. MINOR COMPONENT	H ₂ O, SO ₂			
	TEMPERATURE, °F	CO ₂			
	PRESSURE, PSIG	280-400			
BY-PRODUCTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	1. FUEL GAS				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 17
UNIT OPERATION NUMBER 80
DESCRIPTION AIR SEPARATION & OXIDANT FEEDING

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		AIR BATTERY LIMIT O ₂ , N ₂ Ar AMBIENT ATMOSPHERIC			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		OXYGEN UNIT OP. 20 O ₂ N ₂ , Ar 100 to -298 15-675	NITROGEN VARIOUS N ₂ O ₂ , Ar 100 to -320 0-100		
BY-PRODUCT	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		ARGON SPECIALTY SALES Ar N ₂ AMBIENT AS REQUIRED			
RAW MATERIALS	1. MOLECULAR SIEVES 2. 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 3
UNIT OPERATION NUMBER 81
DESCRIPTION FINAL SOLIDS DISPOSAL

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	WASTE SOLIDS			
	ORIGIN	UNIT OP. 31			
	MAJOR COMPONENTS	ASH, SLAG, LIME			
	SIG. MINOR COMPONENTS	H ₂ O			
	TEMPERATURE, °F	AMBIENT			
	PRESSURE, PSIG	ATMOSPHERIC			
OUTPUTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	DESTINATION				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
BY-PRODUCTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	1. FIXATION CHEMICALS				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 6
UNIT OPERATION NUMBER 82
DESCRIPTION BY-PRODUCT STORAGE & LOADING

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		CRUDE PHENOL UNIT OP. 34 PHENOL H ₂ O AMBIENT ATMOSPHERIC	AMMONIA UNIT OP. 35 NH ₃ --- AS REQUIRED	TAR UNIT OP. 32 TAR, OIL NH ₃ , H ₂ S AMBIENT ATMOSPHERIC	COAL FINES UNIT OP. 11 COAL ASH, H ₂ S AMBIENT ATMOSPHERIC
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		CRUDE PHENOL SALES PHENOL H ₂ O AMBIENT ATMOSPHERIC	AMMONIA SALES NH ₃ --- AS REQUIRED	TAR SALES TAR, OIL NH ₃ , H ₂ S AMBIENT ATMOSPHERIC	COAL FINES SALES COAL H ₂ S, ASH AMBIENT ATMOSPHERIC
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. NATURAL GAS 2. 3. 4. 5. 6. 7. 8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 18
UNIT OPERATION NUMBER 83
DESCRIPTION SULFUR STORAGE AND LOADING

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	SULFUR			
	ORIGIN	UNIT OP. 36			
	MAJOR COMPONENTS	SULFUR			
	SIG. MINOR COMPONENTS	H ₂ S			
	TEMPERATURE, °F	280-320			
	PRESSURE, PSIG	ATMOSPHERIC			
		NO. 1	NO. 2	NO. 3	NO. 4
OUTPUTS	DESCRIPTION	SULFUR	VENT GAS		
	DESTINATION	BATTERY LIMIT	UNIT OP. 41		
	MAJOR COMPONENT	SULFUR	O ₂ , N ₂		
	SIG. MINOR COMPONENT	---	H ₂ S		
	TEMPERATURE, °F	280-320	280-320		
	PRESSURE, PSIG	ATMOSPHERIC	ATMOSPHERIC		
		NO. 1	NO. 2	NO. 3	NO. 4
BY-PRODUCTS	DESCRIPTION	SULFUR			
	USES	SALES			
	MAJOR COMPONENT	SULFUR			
	SIG. MINOR COMPONENT	---			
	TEMPERATURE, °F	280-320			
	PRESSURE, PSIG	ATMOSPHERIC			
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 8
UNIT OPERATION NUMBER 84
DESCRIPTION STEAM GENERATION

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4*
		BFW UNIT OP. 85 H ₂ O DISS. SOLIDS 250 400-1500	IBG UNIT OP. 23 H ₂ , CO, CO ₂ H ₂ O 100-200 600		COAL UNIT OP. 10 COAL ASH, H ₂ O AMBIENT ATMOSPHERIC
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		STEAM BATTERY LIMIT H ₂ O DISS. SOLIDS 450-900 400-1500			FLUE GAS UNIT OP. 86 SO ₂ , H ₂ S H ₂ O, NH ₃ 200-600 ATMOSPHERIC
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. ETHYLENE DIAMINE TETRAACETIC ACID 2. OCTYLDECYLAMINE 3. ANTIFOAM AGENT 4. SODIUM SULFITE 5. HYDRAZINE 6. 7. 8.	<p>* These inputs & output streams relate to coal fired boilers.</p>			

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 7
UNIT OPERATION NUMBER 85
DESCRIPTION RAW WATER TREATMENT

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		RIVER WATER BATTERY LIMIT H ₂ O, O ₂ , SOLIDS IRON, SILICA AMBIENT 30-100			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		FILTERED H ₂ O BATTERY LIMIT H ₂ O DISS. SOLIDS AMBIENT 30-100	BFW UNIT OP. 84 H ₂ O SiO ₂ 250 400-1500	REGEN WASTE UNIT OP 33,31 H ₂ O, SOLIDS H ₂ SO ₄ , NaOH 120 50	LIME SLUDGE UNIT OP. 31 Ca(OH) ₂ , MgO MINERAL SALT AMBIENT 2-10
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. SULFURIC ACID, SODIUM HYDROXIDE 2. LIME, ALUM, CHLORINE 3. POLYELECTROLYTE 4. NaCl, FeCl 5. ION EXCHANGE RESIN 6. REVERSE OSMOSIS & ULTRAFILTRATION ELEMENT 7. COAGULANT AIDS 8. FILTER SAND, ACTIVATED CARBON				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 8

UNIT OPERATION NUMBER 86

DESCRIPTION FLUE GAS TREATMENT

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	FLUE GAS			
	ORIGIN	UNIT OP. 84			
	MAJOR COMPONENTS	SO ₂ , CO ₂ , H ₂ O			
	SIG. MINOR COMPONENTS	N ₂ , O ₂ , H ₂ S			
	TEMPERATURE, °F	300-600			
	PRESSURE, PSIG	1-2			
OUTPUTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	DESTINATION	VENT GAS	WASTE SOLID	PURGE LIQUOR	
	MAJOR COMPONENT	ATMOSPHERE	UNIT OP. 81	UNIT OP. 33	
	SIG. MINOR COMPONENT	H ₂ O, O ₂ , CO ₂ , N ₂	SODIUM SALT	SODIUM SALT	
	TEMPERATURE, °F	SO ₂ , SO ₃	CALCIUM SALT	---	
BY-PRODUCTS	TEMPERATURE, °F	125-150	70-110	150-200	
	PRESSURE, PSIG	ATMOSPHERIC	ATMOSPHERIC	ATMOSPHERIC	
	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	1. LIME				
	2. SODIUM CARBONATE				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER

5

UNIT OPERATION NUMBER

87

DESCRIPTION

PLANT ELECTRICAL SYSTEM

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	NO PROCESS RELATED DATA			
	ORIGIN				
OUTPUTS	MAJOR COMPONENTS				
	SIG. MINOR COMPONENTS				
BY-PRODUCTS	TEMPERATURE, °F				
	PRESSURE, PSIG				
RAW MATERIALS	DESCRIPTION				
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER 1

UNIT OPERATION NUMBER 88

DESCRIPTION BUILDINGS & SUPPORT FACILITIES

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	POTABLE H ₂ O			
	ORIGIN	UNIT OP. 85			
	MAJOR COMPONENTS	H ₂ O			
	SIG. MINOR COMPONENTS	CHLORINE			
	TEMPERATURE, °F	AMBIENT			
	PRESSURE, PSIG	0-20			
		NO. 1	NO. 2	NO. 3	NO. 4
OUTPUTS	DESCRIPTION	INST. AIR	SAN. WASTES		
	DESTINATION	BATTERY LIMIT	UNIT OP. 37		
	MAJOR COMPONENT	AIR	WASTES		
	SIG. MINOR COMPONENT	---	---		
	TEMPERATURE, °F	100	AMBIENT		
	PRESSURE, PSIG	125	ATMOSPHERIC		
		NO. 1	NO. 2	NO. 3	NO. 4
BY-PRODUCTS	DESCRIPTION				
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
		NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1.				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX C
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER

4

UNIT OPERATION NUMBER

89

DESCRIPTION

CONTROL & INSTRUMENTATION

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
INPUTS	DESCRIPTION	NO DATA REPORTED			
	ORIGIN				
MAJOR COMPONENTS					
SIG. MINOR COMPONENTS					
TEMPERATURE, °F					
PRESSURE, PSIG					
OUTPUTS	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	DESTINATION				
BY-PRODUCTS	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
	DESCRIPTION	NO. 1	NO. 2	NO. 3	NO. 4
	USES				
RAW MATERIALS	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
	1.				
	2.				
	3.				
	4.				
5.					
6.					
7.					
8.					

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE EPRI AF-741		1	1	1	1
CASE		1	2	3	4
COAL FEED, TON/DAY		19,680	20,270	20,000	22,783
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		810.736	907.566	867.257	1,067.870
DATE OF MONEY		Mid-76	Mid-76	Mid-76	Mid-76
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER	/204	/213	/209	/226
	MM DOLLARS/YR	4.667	4.873	4.782	5.171
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	19.650	33.193	17.500	34.639
	INITIAL CHARGE				
	MM DOLLARS	3.190	4.786	3.375	6.015
	UTILITY COSTS				
	MM DOLLARS/YR	2.13	2.56	2.18	3.18
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	12.945	14.243	13.746	16.514
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	17.408	19.028	18.499	22.098
	LABOR TO MATERIAL RATIO (NOTE 1)	43/57	43/57	43/57	43/57
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	30.353	33.271	32.245	38.612

NOTES: (1) L/M Ratio was 50/50 for some units; others were 40/60.
L/M shown is calculated.

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 2

REFERENCE EPRI AF-416		2	2	2	2	
CASE		MXL	MXH	MAL	MAH	
COAL FEED, TON/DAY		12,000	12,000	12,000	12,000	
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		68.071	456.009	608.687	416.078	
DATE OF MONEY		Mid-75	Mid-75	Mid-75	Mid-75	
OPERATING COSTS	OPERATING LABOR MANHOURS/MANPOWER MM DOLLARS/YR		8.616	5.768	7.700	5.263
	OPERATING MATERIALS MM DOLLARS/YR					
	CATALYST AND CHEMICALS	ANNUAL USAGE MM DOLLARS/YR	0.548	0.367	0.500	0.335
		INITIAL CHARGE MM DOLLARS	3.916	2.622	3.500	2.392
		UTILITY COSTS MM DOLLARS/YR		0.159	0.225	0.082
	MAINTENANCE LABOR MANHOURS/MANPOWER MM DOLLARS/YR		12.923	8.653	11.550	7.895
	MAINTENANCE MATERIALS MM DOLLARS/YR		18.014	12.061	16.100	11.005
MAINTENANCE COSTS	LABOR TO MATERIAL RATIO		42.58	42.58	42.58	42.58
	TOTAL MAINTENANCE MM DOLLARS/YR		30.937	20.714	27.650	18.900

APPENDIX C
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 2 OF 2

REFERENCE EPRI AF-416 CASE		2 MSL	2 MSH	2 EXL	2 EXH
COAL FEED, TON/DAY		12,000	12,000	12,000	12,000
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		407.216	391.961	425.951	425.852
DATE OF MONEY		Mid-75	Mid-75	Mid-75	Mid-75
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	5.151	4.958	5.388	5.387
	OPERATING MATERIALS				
	MM DOLLARS/YR	52.882	51.564	55.360	53.275
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	0.328	0.316	0.343	0.343
	INITIAL CHARGE				
	MM DOLLARS	2.341	2.254	2.449	2.449
MAINTENANCE COSTS	UTILITY COSTS				
	MM DOLLARS/YR	0.035	0.028	0.043	0.062
	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	7.727	7.437	8.082	8.080
MAINTENANCE COSTS	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	10.771	10.367	11.266	11.264
	LABOR TO MATERIAL RATIO	42/58	42/58	42/58	42/58
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	18.498	17.804	19.348	19.344

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE FE-2447-13 CASE		3 1	3 1-A	3 1-B	3 1-C
COAL FEED, TON/DAY		27,300	27,300	27,400	27,300
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		1,084	991	1,242	1,050
DATE OF MONEY		Oct 77	Oct 77	Oct 77	Oct 77
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER	/261	/241	/257	/259
	MM DOLLARS/YR	6.9	6.9	6.9	6.9
	OPERATING MATERIALS				
	MM DOLLARS/YR	1.7	1.7	1.7	1.7
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	5.100	4.600	6.000	5.100
MAINTENANCE COSTS	INITIAL CHARGE				
	MM DOLLARS	6.000	6.000	8.000	6.000
	UTILITY COSTS				
	MM DOLLARS/YR	1.4	1.4	1.4	1.4
	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	29.52	29.52	29.52	29.52
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	19.68	19.68	19.68	19.68
	LABOR TO MATERIAL RATIO	60/40	60/40	60/40	60/40
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	49.2	49.2	49.2	49.2

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE PB-264-702		4	4		
CASE		K-T	B&W		
COAL FEED, TON/DAY		3,400	NO COST		
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		173.2	DATA		
DATE OF MONEY		4th Quarter 1976			
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER	169,128/			
	MM DOLLARS/YR	1.285			
	OPERATING MATERIALS				
	MM DOLLARS/YR	0.386			
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	0.133			
	INITIAL CHARGE				
	MM DOLLARS				
	UTILITY COSTS				
	MM DOLLARS/YR	6.917			
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	2.598			
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	2.598			
	LABOR TO MATERIAL RATIO	50/50			
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	5.196			

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 2

REFERENCE		EPRI AF-642	S	S	S	S
CASE			MACW	MXSC	EAHC	EXHC
COAL FEED, TON/DAY			10,000	10,000	10,000	10,000
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS			588.725	566.630	542.122	543.800
DATE OF MONEY			Mid-76	Mid-76	Mid-76	Mid-76
OPERATING COSTS	OPERATING LABOR MANHOURS/MANPOWER MM DOLLARS/YR		3.653	2.884	2.590	2.692
	OPERATING MATERIALS MM DOLLARS/YR					
	CATALYST AND CHEMICALS	ANNUAL USAGE MM DOLLARS/YR	0.373	0.344	0.429	0.391
		INITIAL CHARGE MM DOLLARS	1.072	0.800	1.351	0.846
	UTILITY COSTS MM DOLLARS/YR		1.150	1.197	1.179	1.189
MAINTENANCE COSTS	MAINTENANCE LABOR MANHOURS/MANPOWER MM DOLLARS/YR		8.088	7.580	7.531	7.289
	MAINTENANCE MATERIALS MM DOLLARS/YR		12.132	11.370	11.296	10.934
	LABOR TO MATERIAL RATIO		40/60	40/60	40/60	40/60
	TOTAL MAINTENANCE MM DOLLARS/YR		20.220	18.950	18.827	18.223

APPENDIX D

SUMMARY OF FACILITY RELATED COSTS

OPERATIONS AND MAINTENANCE

SHEET 2 OF 2

REFERENCE EPRI AF-642 CASE		5 EALC	5 EXTC	5 EXTC	
COAL FEED, TON/DAY		10,000	10,000	10,000	
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		685.133	619.470	639.632	
DATE OF MONEY		Mid 76	Mid 76	Mid 76	
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	2.596	2.692	2.692	
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	0.344	0.262	0.284	
	INITIAL CHARGE				
	MM DOLLARS	0.528	0.515	0.515	
	UTILITY COSTS				
	MM DOLLARS/YR	1.272	1.354	1.440	
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	9.798	7.882	8.229	
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	14.697	11.822	12.343	
	LABOR TO MATERIAL RATIO	40/60	40/60	40/60	
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	24.495	19.704	20.572	

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 2

REFERENCE EPRI AF-244		6	6	6	6	
CASE		MA	MX	FA	FX	
COAL FEED, TON/DAY		10,000	10,000	10,000	10,000	
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		334.905	408.758	258.531	252.289	
DATE OF MONEY		Mid '75	Mid 75	Mid 75	Mid 75	
OPERATING COSTS	OPERATING LABOR MANHOURS/MANPOWER MM DOLLARS/YR		4.028	4.400	3.284	3.567
	OPERATING MATERIALS MM DOLLARS/YR		1.629	1.127	.800	.575
	CATALYST AND CHEMICALS	ANNUAL USAGE MM DOLLARS/YR	0.314	0.334	0.238	0.165
		INITIAL CHARGE MM DOLLARS	2.345	1.558	2.386	1.299
	UTILITY COSTS MM DOLLARS/YR		0.487	0.543	0.384	0.282
	MAINTENANCE LABOR MANHOURS/MANPOWER MM DOLLARS/YR		6.302	7.655	4.905	4.856
MAINTENANCE COSTS	MAINTENANCE MATERIALS MM DOLLARS/YR		8.788	10.715	6.850	6.869
	LABOR TO MATERIAL RATIO		42/58	42/58	42/58	42/58
	TOTAL MAINTENANCE MM DOLLARS/YR		15.090	18.370	11.755	11.725

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 2 OF 2

REFERENCE EPRI AF-244 CASE		6 EAL	6 EXL		
COAL FEED, TON/DAY		10,000	10,000		
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		202.767	209.729		
DATE OF MONEY		Mid 75	Mid 75		
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	3.244	3.171		
	OPERATING MATERIALS				
	MM DOLLARS/YR	.608	.504		
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	0.198	0.181		
	INITIAL CHARGE				
	MM DOLLARS	0.527	0.527		
	UTILITY COSTS				
	MM DOLLARS/YR	0.273	0.211		
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	4.388	4.325		
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	6.176	6.386		
	LABOR TO MATERIAL RATIO	42/58	40/60		
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	10.564	10.711		

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE EPRI AF-531		7	7		
CASE		1	2		
COAL FEED, TON/DAY		7,850	8,250		
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		121.98	133.338		
DATE OF MONEY		Mid 75	Mid 75		
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER		91,700/	91,700/	
	MM DOLLARS/YR		0.7336	0.7336	
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS	ANNUAL USAGE			
		MM DOLLARS/YR	.081	.055	
MAINTENANCE COSTS	CATALYST AND CHEMICALS	INITIAL CHARGE			
		MM DOLLARS	1.655	.069	
	UTILITY COSTS				
	MM DOLLARS/YR		10.323	8.720	
	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR				
MAINTENANCE COSTS	MAINTENANCE MATERIALS				
	MM DOLLARS/YR				
	LABOR TO MATERIAL RATIO				
	TOTAL MAINTENANCE				
MM DOLLARS/YR			4.2332	4.624	

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE FE 1775-13		8			
CASE		Base			
COAL FEED, TON/DAY		43,700			
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS (NOTE 1)		1,866.547			
DATE OF MONEY		Mid 77			
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER (NOTE 2)	/1468			
	MM DOLLARS/YR	30.04			
	OPERATING MATERIALS				
	MM DOLLARS/YR	16.048			
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	7.250			
	INITIAL CHARGE				
	MM DOLLARS	8.846			
	UTILITY COSTS				
	MM DOLLARS/YR	8.846			
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER (NOTE 3)	/988			
	MM DOLLARS/YR (NOTE 4)	46.131			
	MAINTENANCE MATERIALS (NOTE 5)	53.488			
	MM DOLLARS/YR				
	LABOR TO MATERIAL RATIO	46/54			
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	99.619			

- NOTES: (1) Includes \$MM 231.310 for coal mine
(2) 887 for coal mine
(3) 308 for coal mine
(4) $56/280 \times 10^6$ /yr for coal mine
(5) 19.612×10^6 /yr for coal mine equipment operation included in Maintenance Materials

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE FE 1775-18		9	9		
CASE		1	2		
COAL FEED, TON/DAY		1,800	2,250		
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		83.666	250.451		
DATE OF MONEY		Mid 77	Mid 77		
OPERATING COSTS	OPERATING LABOR (NOTE 1)				
	MANHOURS/MANPOWER		/35	/62	
	MM DOLLARS/YR		0.545	0.965	
	OPERATING MATERIALS				
	MM DOLLARS/YR		0.165	0.290	
	CATALYST AND CHEMICALS	ANNUAL USAGE			
		MM DOLLARS/YR	0.830	0.380	
		INITIAL CHARGE			
MAINTENANCE COSTS	MM DOLLARS		0.364	0.065	
	UTILITY COSTS				
	MM DOLLARS/YR				
	MAINTENANCE LABOR		75	108	
	MANHOURS/MANPOWER				
	MM DOLLARS/YR		1.520	4.545	
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR		3.070	9.195	
	LABOR TO MATERIAL RATIO		33/67	33/67	
	TOTAL MAINTENANCE				
	MM DOLLARS/YR		4.590	13.740	

NOTES: (1) Includes Supervision

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE EPRI AF-916 CASE		10 1	10 4	10 5	
COAL FEED, TON/DAY		10,000	10,000	10,000	
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		634.622	669.223	632.511	
DATE OF MONEY		Mid 76	Mid 76	Mid 76	
OPERATING COSTS	OPERATING LABOR MANHOURS/MANPOWER MM DOLLARS/YR		2.974	2.974	2.974
	OPERATING MATERIALS MM DOLLARS/YR				
	CATALYST AND CHEMICALS	ANNUAL USAGE MM DOLLARS/YR	0.371	0.458	0.249
		INITIAL CHARGE MM DOLLARS	0.530	0.654	0.356
		UTILITY COSTS MM DOLLARS/YR	2.715	2.951	2.701
	MAINTENANCE LABOR MANHOURS/MANPOWER MM DOLLARS/YR		6.487	6.748	6.480
	MAINTENANCE MATERIALS MM DOLLARS/YR		9.731	10.122	9.720
MAINTENANCE COSTS	LABOR TO MATERIAL RATIO		40/60	40/60	40/60
	TOTAL MAINTENANCE MM DOLLARS/YR		16.218	16.870	16.200

A-D-13

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APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE EPRI AF-880 (NOTE 1) CASE		11 1	11 2	11 3	
COAL FEED, TON/DAY		1,100	1,100	1,100	
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		233.980	258.710	266.680	
DATE OF MONEY		April 78	April 78	April 78	
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR				
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR				
	INITIAL CHARGE				
	MM DOLLARS				
	UTILITY COSTS				
	MM DOLLARS/YR				
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR				
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR				
	LABOR TO MATERIAL RATIO				
	TOTAL MAINTENANCE				
	MM DOLLARS/YR				

NOTES: (1) No cost breakdown provided

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE FE 1775-7		12			
CASE		1			
COAL FEED, TON/DAY		30,000			
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS (NOTE 1)		1,376.3			
DATE OF MONEY		4th Qtr 75			
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER (NOTE 2)	/978			
	MM DOLLARS/YR	18.92			
	OPERATING MATERIALS				
	MM DOLLARS/YR	9.712			
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	5.175			
	INITIAL CHARGE				
	MM DOLLARS	10.668			
	UTILITY COSTS				
	MM DOLLARS/YR				
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER (NOTE 3)	/796			
	MM DOLLARS/YR	34.340			
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	34.348			
	LABOR TO MATERIAL RATIO	50/50			
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	68.688			

NOTES: (1) Includes \$MM 164.3 for coal mine
(2) 598 for coal mine
(3) 217 for coal mine

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE PROCON CASE		16 BITUMINOUS	16 LIGNITE			
COAL FEED, TON/DAY		21,321	24,641			
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		770.660	634.393			
DATE OF MONEY		1st Qtr 78	1st Qtr 78			
OPERATING COSTS	OPERATING LABOR (NOTE 1)					
	MANHOURS/MANPOWER		8,760/	8,760/		
	MM DOLLARS/YR		5.326	5.326		
	OPERATING MATERIALS					
	MM DOLLARS/YR (NOTE 2)		1.60	1.60		
	CATALYST AND CHEMICALS	ANNUAL USAGE				
		MM DOLLARS/YR	4.584	4.861		
INITIAL CHARGE						
	MM DOLLARS	10.325	11.383			
UTILITY COSTS						
MM DOLLARS/YR		2.161	1.625			
MAINTENANCE COSTS	MAINTENANCE LABOR					
	MANHOURS/MANPOWER					
	MM DOLLARS/YR					
	MAINTENANCE MATERIALS					
MM DOLLARS/YR						
LABOR TO MATERIAL RATIO						
TOTAL MAINTENANCE						
MM DOLLARS/YR		39.246	32.113			

NOTES: (1) Includes supervision
(2) Operating supplies @ 30% of operating labor

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE FE 2542-10		17			
CASE					
COAL FEED, TON/DAY		29,000			
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		732.836			
DATE OF MONEY		1st Qtr 78			
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER	/276			
	MM DOLLARS/YR	5.288			
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS				
MAINTENANCE COSTS	ANNUAL USAGE				
	MM DOLLARS/YR	15.275			
	INITIAL CHARGE				
	MM DOLLARS	14.338			
	UTILITY COSTS				
	MM DOLLARS/YR	2.328			
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER	/220			
	MM DOLLARS/YR	3.917			
	MAINTENANCE MATERIALS (NOTE 1)				
	MM DOLLARS/YR	25.175			
	LABOR TO MATERIAL RATIO				
	TOTAL MAINTENANCE				
	MM DOLLARS/YR				

NOTES: (1) Includes maintenance contracts

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE EPRI AF-753 CASE		19 ETC	19 EATC		
COAL FEED, TON/DAY		10,000	10,000		
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS		619.470	604.752		
DATE OF MONEY		Mid 76	Mid 76		
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	2.692	2.987		
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS				
	ANNUAL USAGE				
	MM DOLLARS/YR	0.262	0.358		
	INITIAL CHARGE				
	MM DOLLARS	0.515	1.285		
MAINTENANCE COSTS	UTILITY COSTS				
	MM DOLLARS/YR	1.354	1.589		
	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR	7.882	8.480		
MAINTENANCE COSTS	MAINTENANCE MATERIALS				
	MM DOLLARS/YR	11.822	12.721		
	LABOR TO MATERIAL RATIO	40/60	40/60		
	TOTAL MAINTENANCE				
	MM DOLLARS/YR	19.704	21.201		

APPENDIX D
SUMMARY OF FACILITY RELATED COSTS
OPERATIONS AND MAINTENANCE

SHEET 1 OF 1

REFERENCE		EPRI AF-1227	22		
CASE			1		
COAL FEED, TON/DAY			22,250		
TOTAL INSTALLED CAPITAL (TIC) MM DOLLARS			1.270		
DATE OF MONEY			1977		
OPERATING COSTS	OPERATING LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR		7.1		
	OPERATING MATERIALS				
	MM DOLLARS/YR				
	CATALYST AND CHEMICALS	ANNUAL USAGE			
MM DOLLARS/YR		11.4			
INITIAL CHARGE					
	MM DOLLARS		23.4		
UTILITY COSTS					
MM DOLLARS/YR					
MAINTENANCE COSTS	MAINTENANCE LABOR				
	MANHOURS/MANPOWER				
	MM DOLLARS/YR		18.9		
	MAINTENANCE MATERIALS				
	MM DOLLARS/YR		26.8		
LABOR TO MATERIAL RATIO		40/60			
TOTAL MAINTENANCE					
MM DOLLARS/YR		45.7			

APPENDIX E

ALTERNATE PRODUCTS SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

UNIT OPERATION NUMBER		PAGE
90	SHIFT CONVERSION	E-1
91	METHANATION	E-2
92	PRODUCT GAS DRYING	E-3
93	METHANOL SYNTHESIS	E-4
94	GASOLINE SYNTHESIS	E-5
95	HYDROGEN RECOVERY	E-6

APPENDIX E
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER

UNIT OPERATION NUMBER

DESCRIPTION

90

SHIFT CONVERSION

SHEET 1 OF 1

		NO. 1	NO. 2	NO. 3	NO. 4
		FEED GAS	STEAM		
INPUTS	DESCRIPTION	UNIT OP 22	UNIT OP 84		
	ORIGIN	CO, H ₂ , CO ₂ , H ₂ O	H ₂ O		
	MAJOR COMPONENTS	H ₂ S, COS	---		
	SIG. MINOR COMPONENTS	350-400	600		
	TEMPERATURE, °F	400	400-1500		
	PRESSURE, PSIG				
		NO. 1	NO. 2	NO. 3	NO. 4
		SHIFT GAS	SOUR H ₂ O		
OUTPUTS	DESCRIPTION	UNIT OP 91, 92	UNIT OP 33		
	DESTINATION	H ₂ , CO ₂ , CO, H ₂ O	H ₂ O		
	MAJOR COMPONENT	H ₂ S, COS	H ₂ S, COS		
	SIG. MINOR COMPONENT	100	230		
	TEMPERATURE, °F	375	375		
	PRESSURE, PSIG				
		NO. 1	NO. 2	NO. 3	NO. 4
BY-PRODUCTS	DESCRIPTION				
	USES				
	MAJOR COMPONENT				
	SIG. MINOR COMPONENT				
	TEMPERATURE, °F				
	PRESSURE, PSIG				
		NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. HIGH TEMP. CATALYST				
	2. LOW TEMP. CATALYST				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX E
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER _____

UNIT OPERATION NUMBER 91

DESCRIPTION

METHANATION

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		FEED GAS UNIT OP 22 CO, H ₂ H ₂ S, CO ₂ 100			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SNG UNIT OP 93 CH ₄ CO, H ₂ , CO ₂ 100	CONDENSATE UNIT OP 85 H ₂ O CO ₂ 100		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		STEAM MOTIVE PWR H ₂ O - 598 1500			
RAW MATERIALS	1. METHANATION CATALYST - NICKEL OXIDE 2. SULFUR POLISH - ZINC OXIDE 3. 4. 5. 6. 7. 8.				

APPENDIX E
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER

UNIT OPERATION NUMBER

DESCRIPTION

92

PRODUCT GAS DRYING

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SYN. NAT. GAS UNIT OP 92 CH ₄ H ₂ O, H ₂ 100 600			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		DRY SNG PIPELINE CH ₄ CO, H ₂ 100 600			
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. TRIETHYLENE GLYCOL 2. 3. 4. 5. 6. 7. 8.				

APPENDIX E
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER _____

UNIT OPERATION NUMBER 93

DESCRIPTION

METHANOL SYNTHESIS

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		FEED GAS UNIT OP 22 CO, H ₂ CO ₂ , CH ₄ 300+			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		METHANOL BATTERY LIMIT CH ₃ OH H ₂ O 100 ATMOSPHERIC	PURGE GAS BATTERY LIMIT CO ₂ , CH ₄ CO 200 200		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. SYNTHESIS CATALYST - COPPER BASE 2. 3. 4. 5. 6. 7. 8.				

APPENDIX E
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER

UNIT OPERATION NUMBER

DESCRIPTION

94

GASOLINE SYNTHESIS

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		METHANOL UNIT OP 92 CH ₃ OH H ₂ O 100 5			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		GASOLINE BLENDING C ₆ ⁺ --- 100	ALKYLATE BLENDING C ₆ ⁺ --- 100	BUTANES BLENDING C ₄ --- 100	LIGHT GAS PLANT FUEL CH ₄ , C ₂ H ₆ --- 100
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		LPG FUEL, FEED PROPANE BUTANE 100 100-250			
RAW MATERIALS	1. SYNTHESIS CATALYST - ZEOLITE				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

APPENDIX E
SYSTEM STREAM CHARACTERIZATION AND RAW MATERIALS

SYSTEM NUMBER

UNIT OPERATION NUMBER

DESCRIPTION

95

HYDROGEN RECOVERY

SHEET 1 OF 1

INPUTS	DESCRIPTION ORIGIN MAJOR COMPONENTS SIG. MINOR COMPONENTS TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		SHIFT GAS UNIT OP 90 CO, H ₂ , CO ₂ H ₂ O 100 100			
OUTPUTS	DESCRIPTION DESTINATION MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
		HYDROGEN BATTERY LIMIT H ₂ CO 100 ATMOSPHERIC	PURGE GAS		
BY-PRODUCTS	DESCRIPTION USES MAJOR COMPONENT SIG. MINOR COMPONENT TEMPERATURE, °F PRESSURE, PSIG	NO. 1	NO. 2	NO. 3	NO. 4
RAW MATERIALS	1. MOLECULAR SIEVES				
	2.				
	3.				
	4.				
	5.				
	6.				
	7.				
	8.				

**APPENDIX F
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER -
UNIT OPERATION NUMBER 90
DESCRIPTION SHIFT CONVERSION

SHEET 1 OF 1

REFERENCE/CASE	3/1	17/ 3/1	22/1 2/0	13/MYCAS 2/0	13/BIGAS 2/0	13/SYNTNAME 2/0	
NO. OF PARALLEL TRAINS							
OPERATING/SPARE							
SYSTEM COST, MM \$	12.8	51.776	66.0	27.0	29.0	31.0	
DATE OF MONEY	10-77	1st Qtr 78	1977	1/76	1/76	1/76	
COST DRIVERS	1. H ₂ /CO ratio in feed	2.07/1	0.439	1.2	1.02	1.18	
	2. MOLE % S IN FEED	0.00	0.12	0.61	0.95	1.29	
	3. MOLE CO SHIFTED	2824	NA	4456	7142	3862	
	4. PRESSURE, PSIA	415	420	1166	1200	985	
	5. SHIFT FEED MOLE/HR	36,303	39,818	39,328	57,154	42,321	
DESIGN DRIVERS	1. STEAM/DRY GAS RATIO	0.59	0.47	0.67	0.65	0.65	
	2. FEED GAS TEMPERATURE, °F	367	331	462	462	437	
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	NA	10	500	0	0	
	2. H.P. STEAM, Mlb/HR(P SIG)	0	914.9 (550)	-500 (600)	0 BFW HEATED	0 BFW HEATED	
	3. M.P. STEAM, Mlb/HR(P SIG)	0	0	-835 (150)			
	4. L.P. STEAM, Mlb/HR	0	0	0			
	5. BOILER FEEDWATER, GPM	NA	0	0			
	6. COOLING WATER, GPM	NA	0	9905	2478	2911	2889

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APPENDIX P
SYSTEM CHARACTERIZATION

SYSTEM NUMBER

UNIT OPERATION NUMBER 91

DESCRIPTION METHANATION

SHEET 1 OF 1

REFERENCE/CASE NO. OF PARALLEL TRAINS OPERATING/SPARE SYSTEM COST, MM \$ DATE OF MONEY	3/1	17/LARGE 28.792 1st Qtr 78	13/WTGAS	13/8IGAS	13/LARGE	13/SYSTEM
1. MM\$/HR OF, FINISHED 2. FEED GAS TEMPERATURE: 3. 4. 5.	NA	73	5021 60	6056 60	7849 75	4947 60
DESIGN 1. CAPACITY MM SCFD 2. FEED GAS PRESSURE 3. 4. 5.	209.6 NA	778 280	259.8 1117	307.7 1115	343.8 355	252.0 880
SYSTEM INTERFACES 1. ELECTRICITY, KW 2. H.P. STEAM, MIB/HR (PSIG) 3. M.P. STEAM, MIB/HR 4. L.P. STEAM, MIB/HR 5. BOILER FEEDWATER, GPM 6. COOLING WATER, GPM	NA NA NA NA NA NA	615 -1,605 (600) 0 0 3,648 13,600	-517.5 (1500)	0764.6 (1500)	-483.8 (600)	-535.8 (1500)

SYSTEM NUMBER		UNIT OPERATION NUMBER		SNG DRYING		SHEET 1 OF 1	
DESCRIPTION		3/1		17/0			
REFERENCE/CASE		3		3/0			
NO. OF PARALLEL TRAINS		0.5		13.095			
OPERATING/SPARE		10-77		1st Qtr 78			
SYSTEM COST, MM \$		0.11		10.37			
DATE OF MONEY							
1. A H ₂ O IN FEED							
2.							
3.							
4.							
5.							
1. CAPACITY - MMSCFM		5.86		11.25			
2. TEG CIRCULATION - GPM		NA		7			
3.							
4.							
5.							
1. ELECTRICITY, KW		NA		448.7			
2. H.P. STEAM, MIB/HR (PSIG)		NA		256.2 (550)			
3. M.P. STEAM, MIB/HR		NA		0			
4. L.P. STEAM, MIB/HR		NA		0			
5. BOILER FEEDWATER, GPM		NA		0			
6. COOLING WATER, GPM		NA		19,940			

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APPENDIX F
SYSTEM CHARACTERIZATION

SYSTEM NUMBER -
UNIT OPERATION NUMBER 91
DESCRIPTION METHANOL SYNTHESIS

SHEET 1 OF 1

REFERENCE/CASE	3/1	22/1					
NO. OF PARALLEL TRAINS	3						
OPERATING/SPACE		2/0					
SYSTEM COST, MM \$	54.4	112					
DATE OF MONEY	10-77	1977					
COST DRIVERS	1. H ₂ /CO RATIO IN FEED	3.54	2				
	2. FEED GAS PRESSURE, PSIG	340	400				
	3.						
	4.						
	5.						
DESIGN DRIVERS	1. CAPACITY, Mlb/hr	958.0	1,290.8				
	2. FEED GAS PURITY	NA	<1 ppm S				
	3.						
	4.						
	5.						
SYSTEM INTERFACES	1. ELECTRICITY, KW	NA	1,696				
	2. H.P. STEAM, Mlb/HR	NA	0				
	3. H.P. STEAM, Mlb/HR (PRIG)	NA	(-1704.6) (150)				
	4. L.P. STEAM, Mlb/HR	NA	0				
	5. BOILER FEEDWATER, GPM	NA	0				
	6. COOLING WATER, GPM	NA	5,726				

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APPENDIX P
SYSTEM CHARACTERIZATION

SYSTEM NUMBER		UNIT OPERATION NUMBER 94		DESCRIPTION		CASUALTY SYNOPSIS		SHEET 1 OF 1	
REFERENCE/CASE	NO. OF PARALLEL TRAINS	3/1	3						
OPERATING/SPARE	SYSTEM COST, MM \$	69.8							
DATE OF MONEY	DATE OF MONEY	10-77							
1. SYSTEM PRESSURE, PSIG	2. SYSTEM TEMPERATURE	300							
3. COST	4. DRIVERS								
5. DESIGN	1. CAPACITY, MIB/HR	614.9							
2. DRIVERS	2. ELECTRICITY, KW	NA							
3. DESIGN	3. H.P. STEAM, LB/HR	NA							
4. DRIVERS	4. M.P. STEAM, LB/HR	NA							
5. DESIGN	5. L.P. STEAM, LB/HR	NA							
SYSTEM INTERFACES	6. BOILER FEEDWATER, GPM	NA							
	7. COOLING WATER, GPM	NA							

**APPENDIX F
SYSTEM CHARACTERIZATION**

SYSTEM NUMBER -
UNIT OPERATION NUMBER 95
DESCRIPTION HYDROGEN RECOVERY

SHEET 1 OF 1

A-F-6	REFERENCE/CASE		3/1						
	NO. OF PARALLEL TRAINS								
	OPERATING/SPARE		3						
	SYSTEM COST, MM \$		2.6						
	DATE OF MONEY		10-77						
	COST DRIVERS	1. Δ H ₂ IN GAS	35.46						
		2.							
		3.							
		4.							
		5.							
	DESIGN DRIVERS	1. CAPACITY, MSCFD	3895.8						
		2.							
		3.							
		4.							
		5.							
	SYSTEM INTERFACES	1. ELECTRICITY, KW	NA						
		2. H.P. STEAM, lb/HR	NA						
		3. M.P. STEAM, lb/HR	NA						
		4. L.P. STEAM, lb/HR	NA						
		5. BOILER FEEDWATER, GPM	NA						
		6. COOLING WATER, GPM	NA						